

Physical/Chemical Treatment of Blast
Furnace Wastewaters Using Mobile Pilot Units

Rexnord, Inc.
Milwaukee, WI

Prepared for

Industrial Environmental Research Lab.
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16. ABSTRACT The report documents an in-depth pilot-plant investigation of the applicability of advanced waste treatment methods for upgrading ironmaking blast furnace wastewaters to Best Available Technology Economically Achievable (BATEA) levels. Mobile treatment facilities, designed to operate at a flow rate of 19 liters/min (5 gpm), were used. The methods studied included alkaline chlorination, chemical treatment/clarification, filtration (both dual media and magnetic), ozonation, activated carbon, and reverse osmosis, either singly and/or in combination. These technologies removed such residual contaminants from the wastewater as suspended solids, oil and grease, phenol, cyanide, ammonia, sulfide, and dissolved solids. For each treatment investigated, samples and operational data were obtained for later use in assessing and comparing the adequacy of the method. Priority pollutant samples were also collected from each process investigated. The investigation indicated that: alkaline chlorination, ozonation, and reverse osmosis effectively reduced influent contaminants to below BATEA levels (for all three, proper pretreatment would be required); alkaline chlorination was the least-cost alternative, based on expected capital and operating costs; and ozonation had the smallest space requirements of the three most promising technologies investigated.		
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by

R. Osantowski, A. Geinopolos,
J. Kane, and G. Rollinger

Rexnord, Inc.
5103 West Beloit Road
Milwaukee, Wisconsin 53201

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EPA Project Officer: Robert V. Hendriks

Industrial Environmental Research Laboratory
Office of Environmental Engineering and Technology
Research Triangle Park, NC 27711

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SECTION 1

INTRODUCTION

PROJECT PURPOSE AND APPROACH

This project was initiated to provide an evaluation of the effectiveness of existing treatment technology for upgrading steel mill wastewaters to Best Available Technology (BAT) Economically Achievable limits for Blast Furnace Category scrubber wastewaters. The wastewater tested was a blast furnace effluent from an operating steel mill treatment system that met 1977 Effluent Guidelines for Best Practical Control Technology (BPT) Currently Available. This wastewater contained residual concentrations of suspended solids, BOD, oils and greases, phenols, cyanides, fluorides, ammonia compounds, sulfides, and dissolved solids. The in-depth pilot plant study was performed using mobile facilities designed especially for treating steel plant wastes.

Treatment processes evaluated during the study included: alkaline chlorination, chemical treatment, dual media filtration, magnetic filtration, reverse osmosis, ozonation and activated carbon.

The project objective was achieved through the performance of a program consisting of the three phases outlined below:

PHASE I - Bench Scale Investigations of a Blast Furnace Scrubber Blowdown Wastewater.

PHASE II - Design and Fabrication of the Mobile Treatment Facilities to House the Pilot Scale Equipment.

PHASE III - Operation and Evaluation of the Advanced Waste Treatment Pilot Plant Systems at a Blast Furnace Site

The purpose of the first phase (bench-scale work) was to provide information concerning the treatment methods to be studied for the Phase II design and the Phase III pilot plant investigation (operation and evaluation). Of particular interest were such items as the pretreatment requirements, magnitude of operating variables, expected magnitude of treatment efficiency and effluent quality, selection of equipment and media, and pilot plant system design.

The second phase objective (System Design and Fabrication) was to provide a mobile pilot testing system for evaluating several advanced waste treatment

technologies. The portable treatment system developed included the technology needed to remove the residual contaminants from the blast furnace BPT wastewater to the extent that this wastewater was upgraded to meet BAT requirements as proposed in 1974. Schematic representations of the mobile testing systems are shown in Figures 1 and 2. Trailer No. 1 housed the alkaline chlorination, chemical treatment, magnetic filtration and dual media filtration systems. The ozonator, activated carbon and reverse osmosis technologies were located in Trailer No. 2. The mobile system contained a high degree of automation which greatly assisted the operators during the study. All of the treatment technologies were designed to treat a nominal flow of 18.9 l/min (5 gpm). Figure 3 shows two photographs of a typical field setup at a blast furnace site.

The advanced waste treatment methods, both singularly and in combination, which were investigated on a pilot basis in Phase III included those presented below:

	Key
1. FIL + O + CT	AC: activated carbon
2. ACL + CT + AC	ACL: alkaline chlorination
	CT: chemical treatment
3. CT + FIL + RO + O (on ROB)	FIL: filtration-dual media or magnetic
4. CT + FIL + RO + ACL (on ROB)	O: ozonation
	RO: reverse osmosis
	ROB: reverse osmosis brine

A schematic illustration of the process trains investigated for treatment of the blast furnace wastewater is shown in Figure 4.

For each treatment train investigated, samples and operational data were obtained for later use in assessing, evaluating, appraising and comparing the adequacy of the individual advanced waste treatment methods.

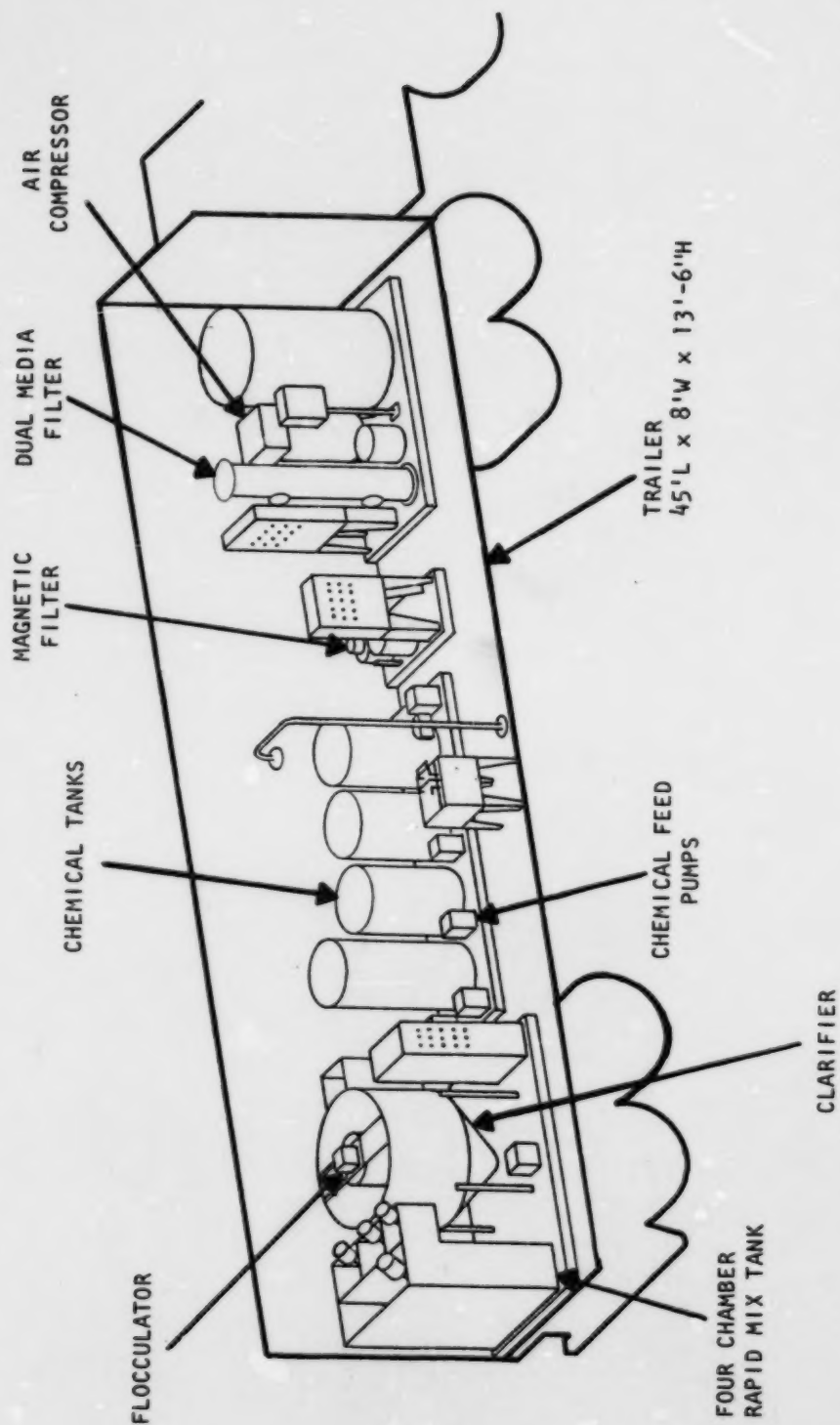
REPORT FORMAT

Evaluation and comparison of the data were performed using the criteria:

1. Process and/or treatment train performance
2. Capital and operating costs
3. Space requirements.

The evaluation resulted in the following benefits:

1. Pilot scale demonstration of the effectiveness of several advanced waste treatment methods for upgrading the steel mill BPT wastewater to meet BAT guidelines.
2. Determination of the basic design and operating characteristics of several advanced waste treatment methods for future application to the steel industry for wastewater discharge.



NOTE: .3048 x ft = m

Figure 1. Steel plant mobile treatment system-trailer No. 1.

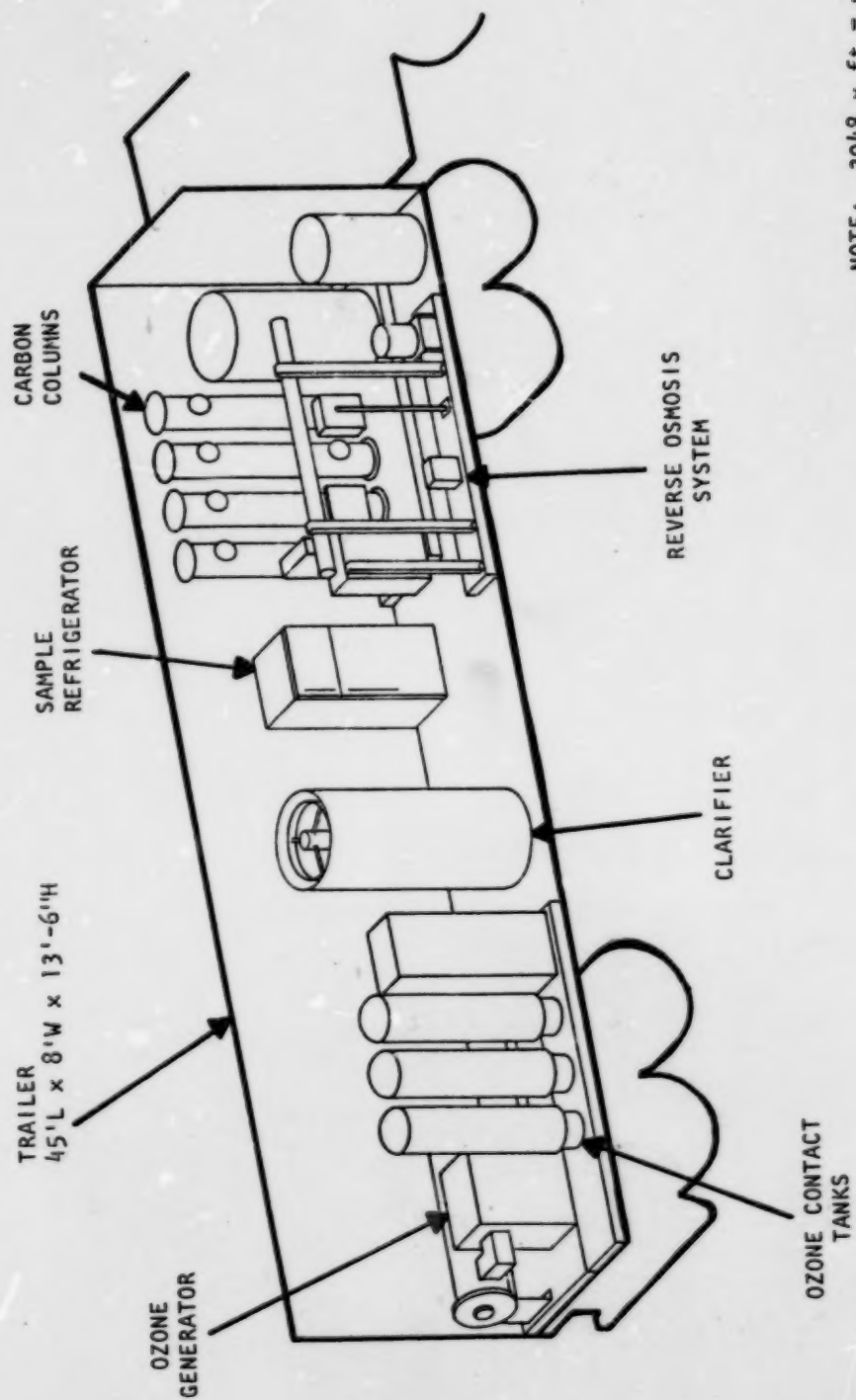


Figure 2. Steel plant mobile treatment system-trailer no. 2.

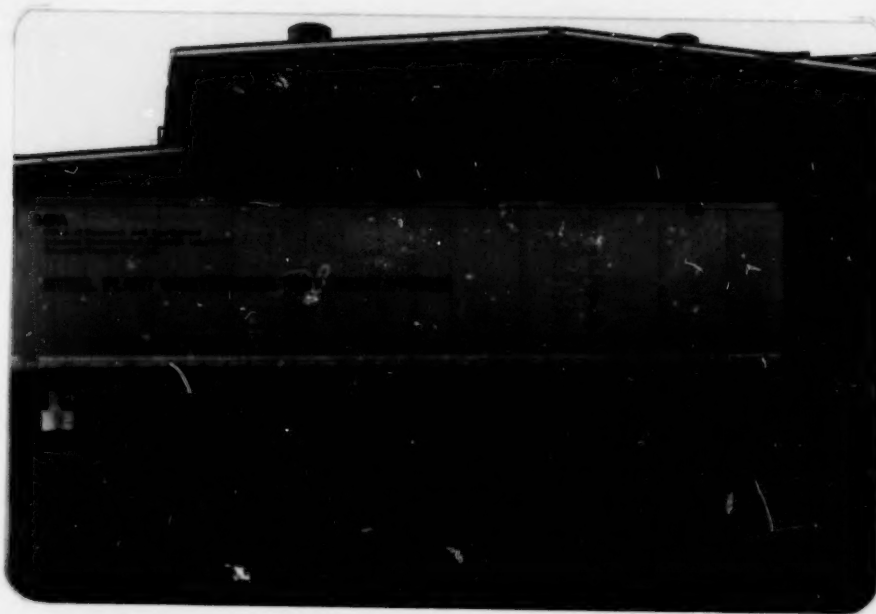
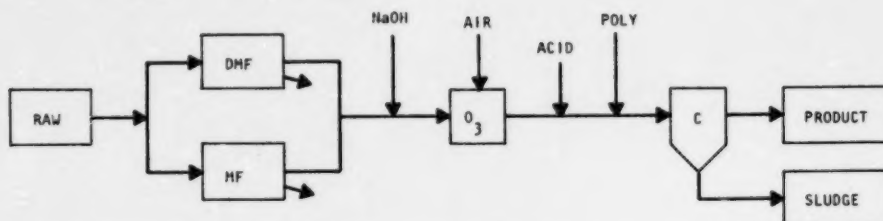
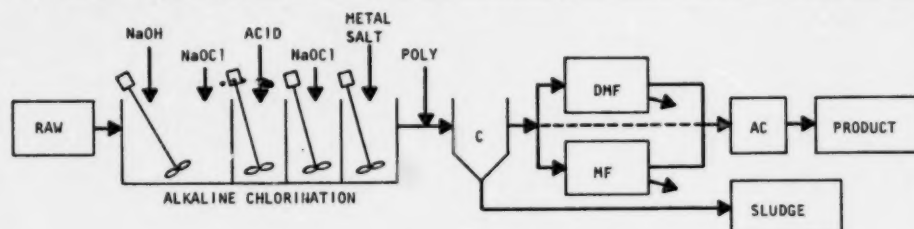


Figure 3. Typical field setup of the mobile treatment facilities - blast furnace test site.

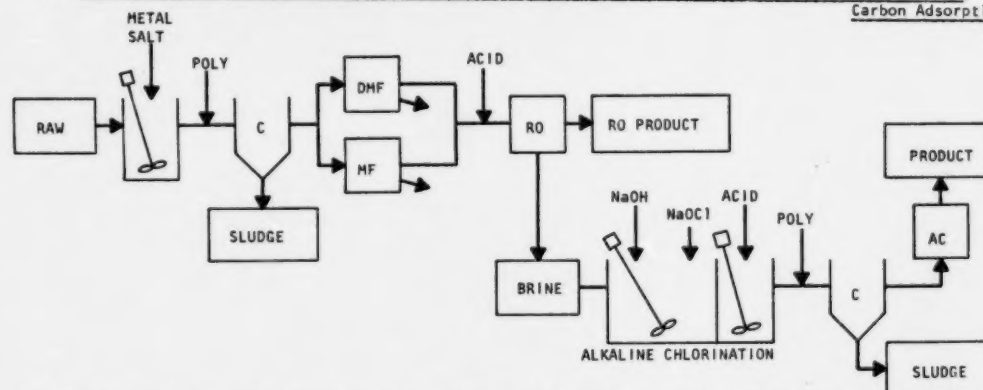
(1) Filtration, Ozonation Clarification



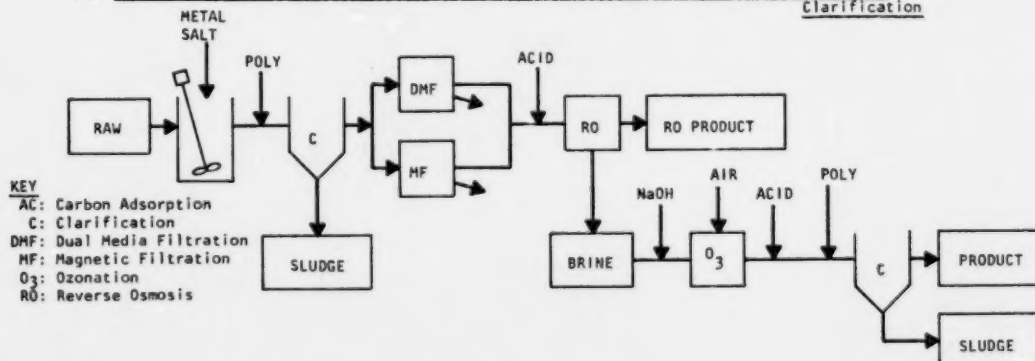
(2) pH Adjustment, Alkaline Chlorination, Chemical Treatment, Clarification, Filtration, Carbon Adsorption



(3) Chemical Treatment, Clarification, Filtration, Reverse Osmosis, Alkaline Chlorination (on brine), Carbon Adsorption



(4) Chemical Treatment, Clarification, Filtration, Reverse Osmosis, Ozonation (On Brine), Clarification



KEY
 AC: Carbon Adsorption
 C: Clarification
 DMF: Dual Media Filtration
 MF: Magnetic Filtration
 O3: Ozonation
 RO: Reverse Osmosis

Figure 4. Process trains investigated for treatment of the blast furnace wastewater.

3. Establishment of the most cost-effective advanced waste treatment processes for producing a BAT effluent suitable for reuse in the mill or for surface discharge.

The sections which follow present a detailed account of the bench scale and pilot scale investigations resulting in the achievement of the project objectives. The report has been divided into individual sections for each of the processes evaluated. Each section discusses the details of the pilot program, presents the results, and concludes with a presentation of cost data. The final section combines the individual processes into treatment schemes capable of meeting the program objective of upgrading the steel plant BPT wastewater to BAT levels. Raw data obtained throughout the study are included in a separate Appendix. A glossary of terms and phrases used throughout the report is included at the end of this report.

SECTION 2

CONCLUSIONS

This report is concerned with the investigation and evaluation of the effectiveness of selected physical-chemical treatment technology on Blast Furnace category scrubber wastewaters. Effectiveness was measured using the criteria: performance, costs and space requirements. Physical-chemical treatment technology investigated included: chlorination, ozonation, reverse osmosis, chemical treatment, clarification, activated carbon and filtration.

1. Performance

The results of the pilot program have demonstrated that the physical-chemical technology investigated was effective in reducing influent blast furnace scrubber wastewater contaminants to below Best Available Technology (BAT) Economically Achievable levels, which were proposed in 1974*.

- a. The most promising methods for removal of substantial amounts of the objectionable contaminants and meeting BAT limitations for blast furnace scrubber wastewaters were chlorination, ozonation and reverse osmosis.
- b. Chemical treatment, clarification, activated carbon and filtration, although not as effective in removing large amounts of the objectionable pollutants, were essential as adjuncts to the primary treatment methods listed in item 1a, above for pretreatment and/or post-treatment purposes.
 - (1) Pretreatment requirements include:
 - (a) For alkaline chlorination: none
 - (b) For reverse osmosis: chemical clarification and filtration
 - (c) For ozonation: filtration
 - (2) Post-treatment requirements:
 - (a) For alkaline chlorination: chemical clarification (and activated carbon where dechlorination is required prior to discharge).

* The BAT limits used hereinafter throughout this report refer to those proposed in 1974. New BAT limits for the Iron and Steel Industry are expected to be proposed in 1980.

(b) For reverse osmosis: brine treatment by alkaline chlorination or ozonation

(1) Following alkaline chlorination of the RO brine, the wastewater would require clarification with polymer (and dechlorination by activated carbon, where required prior to discharge).

(2) After RO brine treatment by ozonation, clarification with polymer is required.

c. The performances of the dual media and magnetic filters were investigated and compared. Based upon suspended solids and turbidity removal data, it was determined that the dual media filter consistently provided a better quality effluent.

2. Costs

The capital and operating costs for the treatment trains involving the primary treatment processes (alkaline chlorination, ozonation and reverse osmosis) were evaluated and compared.

a. Capital Costs

Alkaline chlorination was the least cost alternative treatment train investigated.

By far, the highest cost alternative treatment train investigated was that involving ozonation.

b. Operating Costs

Similarly, alkaline chlorination was the least cost alternative treatment train investigated.

The operating costs for the treatment trains involving ozonation and reverse osmosis were appreciably greater than those for alkaline chlorination.

3. Space Requirements

Evaluation and comparison of the space requirements for the treatment trains involving the three primary treatment processes were also performed.

It was indicated that significantly less space was required for a treatment train utilizing ozonation than for treatment trains involving alkaline chlorination or reverse osmosis.

SECTION 3

RECOMMENDATIONS

1. The results of this study indicated that alkaline chlorination was the cost-effective alternative treatment train investigated in the treatment of blast furnace scrubber blowdown. The results were obtained on a pilot plant basis [18.9 l/min (5 gpm)] over a relatively short test duration (4 weeks).

More comprehensive research data would be available from the testing of a full-scale treatment system over a relatively long duration.

2. The raw wastewater from the blast furnace site selected was very low in objectionable pollutants. Further work should be performed on a blast furnace wastewater containing higher pollutant levels.

SECTION 4

DESCRIPTION OF THE RESEARCH SITE INVESTIGATED

BLAST FURNACE PLANT

General

The blast furnace research site was Bethlehem Steel Corporation, Burns Harbor Plant, located on the southern shore of Lake Michigan. The plant began production in 1964 and the hot metal facilities began production in 1969. The mill utilizes two furnaces: D furnace, with a 10.7 m (35 foot) hearth diameter, was blown-in in December, 1969; C furnace, with a 11.3 m (37 foot) hearth diameter, was blown-in in February, 1972. The combined production of the two furnaces during the pilot study averaged 10,920 metric tons (12,029 tons) of iron in a 24 hour period.

Wastewater Treatment System

Each furnace has its own gas cleaning and cooling system and thickener. Each system consists of a dust catcher, a primary venturi scrubber, a primary washer, a secondary venturi scrubber, a gas cooler, and a thickener. A schematic of the system is presented in Figure 5. Both Eimco thickeners are 27.4 m (90 ft) in diameter, 5 m (16.5 ft) deep with a design overflow rate of 60.3 l/min/m^2 (1.48 gpm/ft^2). Excess quench water from the blast furnace slag quenching is also pumped to each thickener. The thickeners were designed to yield an average effluent suspended solids of 50 mg/l.

The thickener underflows are pumped to the Eimco vacuum filters (2 operating, 1 spare). Filter cake is sent to the Sinter Plant for processing. The clarified overflow from the thickeners flows into the hot well of the Blast Furnace Closed Water Pumping Station. Makeup lake water may be added automatically as required for level control. A liquid scale inhibitor is also added at this point. The water is pumped from the hot well to the five, spray filled cooling towers. Part of the cooling tower system is shown in Figure 6.

The five units operate in series. The cooled water flows by gravity through a bar screen type trash rack to a cold well. Two stage, fixed speed, high lift pumps move the water to the booster pump stations located near each blast furnace. The water is then used for washing and cooling the blast furnace gas. Water is blown down from the discharge side of the high lift pumps as required. Blowdown is based on level control. During

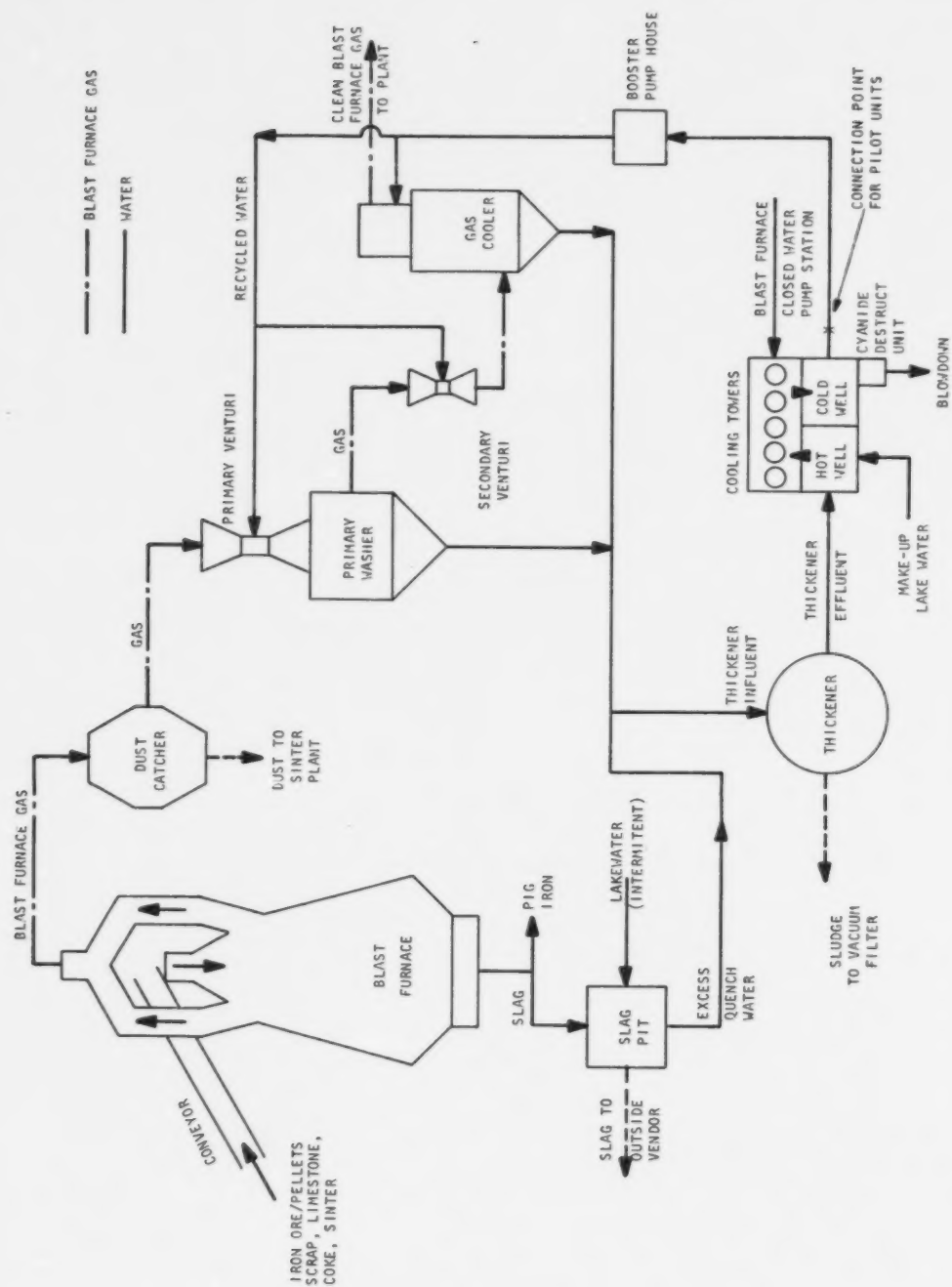


Figure 5. Blast furnace gas cleaning water recirculation system.



Figure 6. Blast furnace cooling tower system.

the pilot study, the gas cleaning water recirculation system blowdown averaged 5,397 m³/day (1.426 MGD). A cyanide destruct unit is located on the blowdown line. The unit consists of an alkaline-chlorination system which is operated on an "as needed" basis to destruct excessive concentrations of cyanide and ammonia in the blowdown. The mobile pilot plant units' feedwater was taken from the return line following blowdown. The blowdown stream has essentially the same characteristics as the return stream due to the addition of makeup water in the hot well. Therefore, the pilot units' feedwater is referred as blowdown in this report.

Best Practicable Control Technology (BPT) Currently Available and Best Available Technology (BAT) Economically Achievable limitations guidelines for this type of plant [Subpart D - Blast Furnace (Iron) Subcategory] are shown on the following page.*

The effluent limitations are based on kilograms (pounds) of pollutant discharged per 1,000 kg (1,000 lb) of iron produced.

Duration of the Study

The period of time spent on-site was approximately 14.5 weeks, including startup time, treatment train switch over and the shutdown of the vans.

* The BAT limits are those proposed in 1974.

Effluent characteristic	Best Control Technology Currently Available		Test Available Technology Economically Achievable	
	Average	Maximum	Average	Maximum
	kg/1000 kg of product		kg/1000 kg of product	
TSS	0.0260	0.0780	0.0130	0.0390
Cyanide A	0.0078	0.0234	0.00013	0.0004
Phenol	0.0021	0.0063	0.00026	0.0008
Ammonia	0.0651	0.1953	0.0052	0.0156
Sulfide			0.00016	0.0005
Fluoride			0.0104	0.0312
pH	Within the range 6.0-9.0		Within the range 6.0-9.0	

The mobile treatment units arrived at the research site on December 11, 1978. Actual testing began on December 15, 1978 and continued through March 30, 1979. The units were shut down from December 23, 1978 through January 1, 1979 for the holidays. The various treatment trains were operational at specific times during the test period. The field testing schedule is presented in Figure 7.

It should be noted that during the pilot testing program, the effluent from the blast furnace blowdown was relatively stable. There was a noticeable change in the wastewater characteristics for a one week period in mid-February, 1979. One of the blast furnace thickeners was taken out of service for this period causing significant increase in some of the parameters being tested. The blast furnace plant including the gas washing system, operates continuously throughout the year.

TREATMENT TRAIN	12/11-12/23/78	12/23-1/1/79	1/2-1/14	1/15-1/28	1/29-2/11	2/12-2/25	2/26-3/11	3/12-3/25	3/26-4/8
START-UP AND SWITCH OVER	■		■	■		■		■	
DMF	■	■							
Mag	■	■	■						
CT + C + Mag								■	
DMF + O + CT + C			■						
Alk. + CT + C + AC				■					
CT + C + DMF + RO + O(On B)					■				
CT + C + DMF + RO + Alk. (On B)							■		
SHUTDOWN	■					■			■

KEY

DMF - Dual Media Filter
 Mag - Magnetic Filter
 CT - Chemical Treatment
 C - Clarification

O - Ozonation
 Alk. - Alkaline Chlorination
 RO - Reverse Osmosis
 B - Reverse Osmosis Brine

Figure 7. 1978-79 field testing schedule blast furnace research site.

SECTION 5

BENCH SCALE TESTING

SUMMARY

Laboratory bench scale tests were performed on a blast furnace wastewater effluent from March 15-30, 1978. The testing was performed in the Rexnord Environmental Research Center laboratory. The purpose of the study was to investigate and evaluate the effectiveness of various advanced treatment methods for meeting Best Available Technology (BAT) Economically Achievable limitations. The wastewater treatment system from which the samples were taken is considered to be meeting Best Practical Control Technology (BPT) Currently Available effluent guidelines. The information obtained from the study was used to obtain design data and the range of operating parameters for the pilot test program.

Processes investigated in the bench scale study included chemical treatment, dual media filtration, activated carbon, reverse osmosis, ozonation and alkaline chlorination.

The results of the bench scale investigation showed that the most promising treatment methods for removal of substantial amounts of the objectionable contaminants and meeting BAT limitations are chlorination, ozonation, activated carbon and reverse osmosis.

Chemical treatment, clarification and filtration, although not effective in removing large amounts of the objectionable pollutants, were essential as adjuncts to the most promising treatment methods listed above.

Alternate treatment trains were recommended for pilot plant investigation and detailed data were obtained for the design and operation of the individual process elements comprising the selected treatment trains.

CONCLUSIONS AND RECOMMENDATIONS

1. Bench scale testing was performed to investigate the effectiveness of various advance treatment methods (filtration, chemical treatment, chlorination, carbon adsorption, ozonation, and reverse osmosis) for meeting BAT effluent requirements. It was concluded that:

- a. The most promising treatment methods for removal of substantial amounts of the objectionable contaminants and meeting BAT limitations are chlorination, ozonation, activated carbon and reverse osmosis.
- b. Chemical treatment, clarification and filtration, although not as effective in removing large amounts of the objectionable pollutants, were essential as adjuncts to the treatment methods listed in Item 1(a), above, for pretreatment and/or post-treatment purposes.

Detailed conclusions and recommendations regarding the above treatment methods are presented below.

2. Reverse Osmosis (RO)

It was concluded that pretreatment using chemical treatment, clarification and filtration was necessary prior to reverse osmosis treatment. RO brine may be treated either by ozonation or alkaline chlorination.

Chemical Treatment

- a. Chemical treatment will be required to reduce the wastewater turbidity to acceptable levels for reverse osmosis application. The use of ferric chloride at dosages of about 50 mg/l and a Betz 1100 polymer dosage of 0.5 mg/l will provide the desired effluent quality. A less suitable alternative would be the use of alum (50-150 mg/l) and Percol 730 polymer (0.75 mg/l). It is recommended that the chemical concentrations indicated be used to plan initial tests and that optimum chemical dosages be determined during the pilot scale study.
- b. The recommended chemical treatment train consists of metal salt addition, 10 second rapid mix, polymer addition, and a 2.5 minute flocculation period. A flocculator-clarifier will be suitable for this treatment. A detention time of 30 to 60 minutes should be used in the pilot tests. The clarifier should be provided with scum removal equipment.

Dual Media Filtration

- c. The use of a 1.0-1.1 mm anthracite coal is recommended for use in the dual media filter pilot study.
- d. Using chemically treated waste as feed, the filter media was capable of removing about two-thirds of the suspended solids and 90 percent of the turbidity. The wastewater pH was unchanged.
- e. In spite of clarification with metal salts and dual media filtration, the membrane fouling potential of the wastewater

remained too high to permit the use of the hollow-fiber reverse osmosis membrane. Also, further consideration will not be given to the tubular configuration due to low membrane surface to unit volume ratio and high operating expense.

- f. It is recommended that the spiral wound membrane be tested during the pilot scale study. The feed stream should consist of chemically pretreated, dual media filtered effluent. A recovery of 75% is feasible without formation of precipitates in the brine.
- g. Reverse osmosis resulted in substantial reductions of all parameters tested except BOD, which was present in low concentration (10 mg/l).

3. Ozonation

It was concluded that ozonation treatment should generally be preceded by filtration and followed by clarification.

- a. Free cyanide and phenol concentrations in filtered wastewater were reduced to acceptable levels at low ozone dosages (about 11 mg/l) without adjustment of pH. Ammonia was not removed under these conditions. Suspended solids concentrations remained low.
- b. High dosages of ozone (about 200 mg/l) were required to reduce the ammonia concentration of filtered wastewater at pH 10 to less than 10 mg/l. Kjeldahl nitrogen, TOC, BOD and soluble manganese concentrations were reduced as well. Treatment at pH 10 results in the formation of a precipitate which can be removed by settling. The use of 0.5 mg/l Betz 1100 polymer increases the rate of solids settling.
- c. Ozonation of reverse osmosis brine at pH 10 was effective in removing phenol at the lowest ozone dosages tested (57 mg/l). The addition of sodium hydroxide was required during ozonation to maintain the pH at the level required for ammonia removal. A dosage of 812 mg/l ozone was required to reduce the ammonia concentration to 13 mg/l. A higher dosage would be needed to reach the 10 mg/l level. Acceptable levels of cyanide were not reached probably because ferricyanide was used to spike the brine instead of potassium cyanide.
- d. Ozonation of the reverse osmosis brine followed by neutralization also produced a precipitate which can be easily settled. Polymer addition (0.5 mg/l Betz 1100) enhances the settling rate and improves effluent suspended solids concentrations.

- e. Based on these tests, it may be difficult to reach acceptable levels of cyanide by ozonation if significant amounts of ferricyanide are present in the waste. Removal of free cyanide can be easily accomplished by ozonation.

4. Alkaline Chlorination

It was concluded that alkaline chlorination, in general, should be preceded by pH adjustment and followed by chemical treatment, clarification and activated carbon (where dechlorination is required prior to discharge).

- a. Alkaline chlorination reduced the ammonia concentration of the wastewater to less than 10 mg/l when a chlorine dose 1.3 times the theoretical amount needed was used (8.7 mg Cl_2 /mg NH_3N). Cyanide concentrations were reduced, but not to the desired level (0.25 mg/l). For this sample, 240 mg/l NaOH and 170 mg/l chlorine were used. After chlorination, 307 mg/l HCl was needed to reduce the pH to 7.4.
- b. The precipitate formed during chlorination was removed by chemical treatment and settling. Treatment with 50 mg/l FeCl and 0.5 mg/l Betz 1100 resulted in a floc having a settling rate of 0.061 m/min (0.2 fpm). A 30 second mixing time and a 2.5 minute flocculation period was utilized. After 30 minutes, the settled sludge volume was less than 10 ml/l.
- c. Treatment of the chlorinated and clarified wastewater with activated carbon resulted in the removal of residual chlorine, the reduction of total cyanide to <0.1 mg/l and a substantial reduction in TOC concentration. Carbon isotherm tests indicated that regenerated bituminous carbon has the following capacities for the waste tested: 62 mg chlorine/g carbon, 1.5 mg CN/g carbon and 18 mg TOC/g carbon. Activated carbon did not significantly reduce the concentration of ammonia remaining after chlorination.
- d. Activated carbon was also effective in reducing the cyanide concentration of unchlorinated wastewater to <0.1 mg/l. The ammonia concentration was not significantly reduced. The capacity of the carbon for adsorbing cyanide from the unchlorinated wastewater was about half the amount measured for adsorption after chlorination.
- e. Alkaline chlorination of reverse osmosis brine required a chlorine dose of 500 mg/l (7.6 mg Cl_2 /mg $\text{NH}_3\text{-N}$) to reduce the ammonia concentration to less than 10 mg/l. Phenol was also destroyed, but total cyanide remained above the desired level.

A detailed discussion of the bench scale findings is presented in Appendix A-1.

SECTION 6

TREATMENT TECHNOLOGIES INVESTIGATED

INTRODUCTION

This section describes the treatment technologies studied in this pilot scale program at the blast furnace research site. For each technology investigated (alkaline chlorination, chemical treatment, magnetic filtration, dual media filtration, reverse osmosis, activated carbon, and ozonation) a general section has been provided for the process introduction, the description of the pilot equipment, and the experimental program performed. A separate discussion on the sampling and analytical procedures used in the study is presented below.

SAMPLING AND ANALYTICAL PROCEDURES

Samples taken during the study periods consisted of grab and 24 hour composites. Grab samples were taken for parameters such as oil and grease and for specified events like chlorine profiles of the activated carbon columns. Composite samples were obtained by collecting hourly discrete samples using ISCO automatic samplers. The sampling procedure and data collection points for each of the treatment processes is shown in Table 1.

The major analytical parameters tested for during the study are presented in Table 2. It should be noted that all of the indicated parameters were not run on all of the samples. Typically, the raw waste was sampled daily throughout the study period. The various product and by-product streams of the different treatment trains were sampled over selected periods. A more detailed description of the sampling and analytical schedule for each process is presented in the Data Evaluation section.

Throughout the sampling program the samples were iced during collection and shipment. Pertinent preservation methods in accordance with EPA (1) were performed in the field immediately after collection. Samples for Biochemical Oxygen Demand (BOD) were split and submitted to a local laboratory. Quality assurance samples were also sent to the lab performing the BOD tests. The remaining samples were packaged, iced, and sent to the Rexnord Inc. laboratory in Milwaukee, Wisconsin. Chances of sample mis-labeling were eliminated by use of unique sample identification numbers, a field sample log and the Rexnord analytical laboratory sample log. A data sheet also accompanied each shipment (See Figure 8). This also provided a chain-of-custody record. The Rexnord laboratory also maintains a quality assurance

TABLE 1. SAMPLING PROCEDURE
AND DATA COLLECTION POINTS

Process	Sampling point/sample type				
	Influent composite	Effluent composite	Influent grab	Effluent grab	Specific grab
Raw waste	X		X		
Alk. chlor.	X		X	X	X
Clarification		X	X	X	
Multi-media filtration backwash	X	X			X
Magnetic filtration backwash	X	X			X
Activated carbon col. #1 col. #2	X	X	X	X	X X
Reverse osmosis product Brine cleaning	X	X X	X	X X	X
Ozone	X	X	X	X	X

Note: (X) indicates sampling point.

TABLE 2. PERTINENT ANALYTICAL PARAMETERS

Analysis	Raw waste	Cyanide treatment	Ammonia treatment	Clarifier inf. eff.	Multi-media		Magnetic filter	
					effluent	backwash	effluent	backwash
pH	X	X	X	X	X		X	
Suspended solids	X	X	X	X	X	X	X	X
Turbidity	X				X			
Oil & grease	X			X	X		X	
TOC	X			X	X		X	
BOD	X			X	X		X	
Cyanide	X	X	X	X	X		X	
Phenol	X			X	X		X	
Color	X				X		X	
Sulfate	X				X		X	
Sulfide	X			X	X		X	
Fluoride	X			X	X		X	
Total dissolved solids	X				X		X	
Alkalinity	X				X		X	
Calcium	X				X		X	
Anions	X				X		X	
Cations	X				X		X	
TKN	X		X		X		X	
NH ₃	X	X	X	X	X		X	
NO ₃ -N	X	X	X	X	X		X	
Chloride	X				X		X	
Cl ₂ residual	X	X	X	X				

Notes: 1) All indicated parameters were not run on all samples.

(continued)

TABLE 2. (continued)

Analysis	Effluent	Act. carbon Profiles	Backwash	Reverse osmosis			Ozone
				Product	Brine	Cleaning	
pH	X	X		X	X	X	X
Suspended solids	X		X	X	X		X
Turbidity	X			X	X		X
Oil & grease	X			X	X		X
TOC	X			X	X		X
BOD	X			X	X		X
Cyanide	X			X	X		X
Phenol	X			X	X		X
Color	X			X	X		X
Sulfate				X			
Sulfide	X			X	X		X
Fluoride	X			X	X		X
Total dissolved solids	X			X	X	X	
Alkalinity				X		X	
Calcium				X	X	X	
Anions				X	X	X	
Cations				X	X	X	
TKN				X	X	X	X
NH ₃	X			X	X		X
NO ₃ -N	X			X	X		X
Chloride	X			X	X		
Cl ₂ residual	X				X		

Notes: 1) All indicated parameters were not run on all samples.

program. All data collected during the study were entered into permanently bound notebooks.

All analytical procedures used during the study period were referenced from Standard Methods for the Examination of Water and Wastewater(2). The equipment used and the specific reference for all analytical procedures is listed below.

1. pH - pH meter, Standard Methods, p. 460.
2. Turbidity - Hach model 2100A turbidimeter, Standard Methods, p. 132.
3. Color - Hellige Aquatester with color comparison discs calibrated at 0, 5, 10, 20, 30, 40, 50, 60 and 70 APHA color units. Waste samples were filtered and diluted with distilled water (if necessary) to reduce color to less than 70 APHA units. When the color of the sample was clearly between that of two color discs, the values for both color discs were recorded. For example, the term 0/5 means that the color of the sample was more than 0 APHA units, but less than 5 APHA units. For plotting and calculation purposes, the value mid-way between the two color discs was used. Standard Methods, p. 64.
4. BOD - Standard Methods, p. 543.
5. Alkalinity - Standard Methods, p. 278.
6. Total Organic Carbon (TOC) - Beckman Model 915 TOC Analyzer, Standard Methods, p. 532.
7. Suspended and Volatile Suspended Solids - Standard Methods, p. 94 and 96.
8. Oil and Grease - Standard Methods, p. 515.
9. Total Phosphate - Standard Methods, p. 476 and 624.
10. Total Dissolved Solids - Standard Methods, p. 93.
11. Ca and Total Hardness - Standard Methods, p. 202.
12. Sulfate - Standard Methods, p. 496.
13. Chloride - Standard Methods, p. 304.
14. Conductivity - Standard Methods, p. 71.
15. Nitrate Nitrogen - Standard Methods, p. 437.
16. Metals - Na, Si, K, Mn, Fe, Al, Ca, Mg - Standard Methods, p. 144.
17. Chemical Oxygen Demand - Standard Methods, p. 550.
18. Ammonia Nitrogen - Standard Methods, p. 407.
19. Kjeldahl Nitrogen - Standard Methods, p. 437.

All pertinent preservation and storage steps, as given in Standard Methods, were followed.

ALKALINE CHLORINATION

Introduction

Alkaline chlorination is an effective treatment process for reducing influent concentrations of cyanide and ammonia. Destruction of cyanide by chlorine oxidation occurs according to the following reaction:



The cyanide is first converted into cyanate at an elevated pH. Chlorine oxidation continues in the second step (neutral pH) with cyanate being oxidized to nitrogen and carbon dioxide, the latter reacting with the alkalinity in the water to form bicarbonates.

Ammonia nitrogen is also converted to nitrogen gas and end products at neutral pH by breakpoint chlorination. The overall reaction may be expressed by the equations shown below:



The alkaline chlorination pilot studies conducted in this project were performed on two wastes; (1) raw wastewater; and (2) reverse osmosis brine.

Description of the Pilot Equipment

The pilot alkaline chlorination system consisted of the reaction tanks, stock tanks, pH and ORP meters, chemical pumps, control panel, etc. necessary to operate the units. There are two separate alkaline chlorination units provided. The first unit is designed to treat influent wastewater flows up to 45.4 l/min (12 gpm). The system consists of four cells:

Cell No. 1	-	1,362.6 l (360 gal.)	[Bypassed]
Cell No. 2	-	227 l (60 gal.)	
Cell No. 3	-	227 l (60 gal.)	
Cell No. 4	-	45.4 l (12 gal.)	

A picture of this system before installation in the research trailers is shown in Figure 9.

The second system was designed to treat reverse osmosis brine at a design flow of 3.785 l/min (1 gpm). Three reaction chambers are provided:

Cell No. 1	-	113.6 l (30 gal.)
Cell No. 2	-	19 l (5 gal.)
Cell No. 3	-	19 l (5 gal.)

This unit is shown in Figure 10. A schematic of a typical system set-up is shown in Figure 11.

Experimental Program

The objectives of the pilot alkaline chlorination tests were to obtain sufficient information for full-scale design. Of vital interest were performance characteristics for determining effluent quality and design parameters to insure the most cost-effective equipment size. Experimental

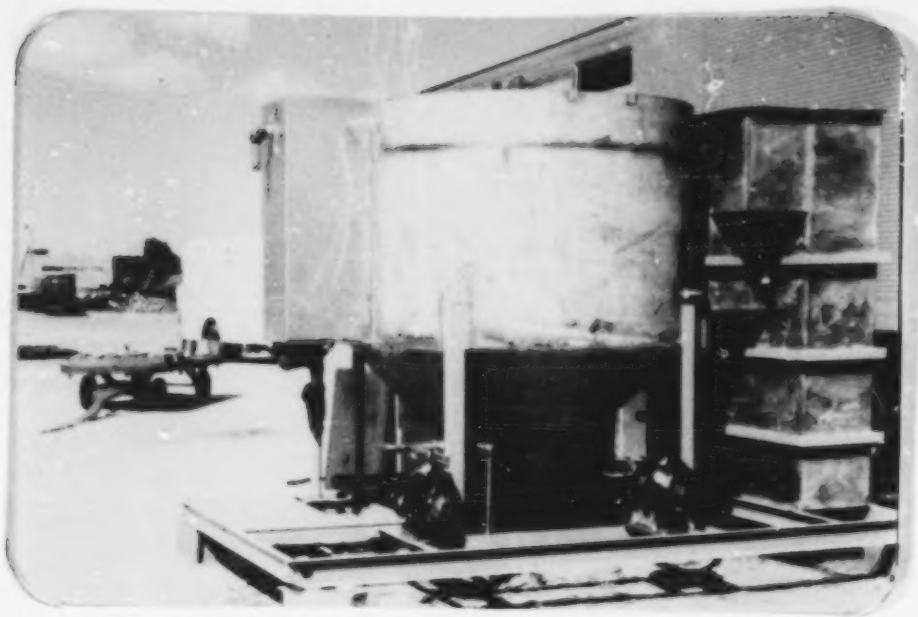


Figure 9. Alkaline chlorination/clarification system for treatment of raw wastewater.



Figure 10. Alkaline chlorination system for treating reverse osmosis brine.

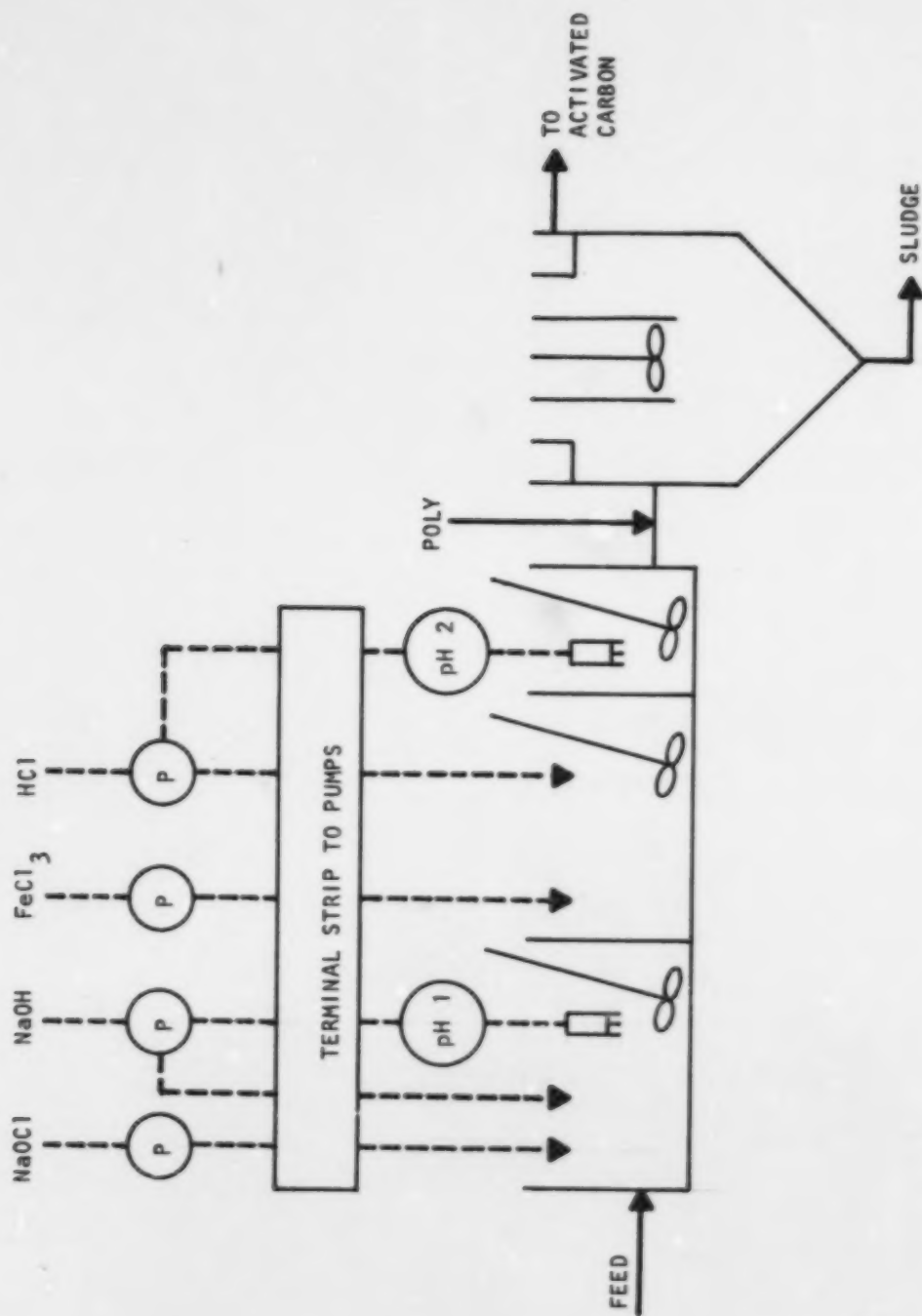


Figure 11. Alkaline chlorination system schematic.

programs were developed for both the raw wastewater and reverse osmosis brine.

Performance characteristics include effluent quality and the consistency of effluent during varying feed conditions. The chlorine to ammonia ratio, contact time, pH, chemical dosage, power requirements, and required mixing are examples of design parameters which can be obtained from pilot testing.

The data obtained during each alkaline chlorination run were recorded on a standard data sheet. Routine recordings included time, flow rate, chemical usage, pH, temperature, and ammonia concentration in each of the reaction tanks. A sample data sheet is shown in Figure 12.

Test Conditions--

Based on prior laboratory bench tests and accepted practice, the following test conditions were selected.

1. Flow Rate: The flow rate was varied from 11.4 l/min (3 gpm) to 32.2 l/min (8.5 gpm) for pilot testing of the raw wastewater. Corresponding detention times were from 7 to 20 minutes in the cyanide removal tank and from 93 to 264 minutes in the ammonia removal system (including clarification volume).

For testing of the reverse osmosis brine, the flow was adjusted from 3.785 l/min (1 gpm) to 22.7 l/min (6 gpm). This allowed a detention time of the cyanide removal tank from 5 to 30 minutes. Detention times for ammonia removal varied from 61 to 366 minutes (including clarification volume).

2. pH Adjustment: The feed water (either raw wastewater or reverse osmosis brine) pH was elevated to 10.0-11.0 in the cyanide tank and then dropped to 6.5-7.0 in the ammonia removal tank. The pH was held relatively constant by the use of pH controllers. The pH probes were checked for calibration periodically.
3. Feed Characteristics: During the study, alkaline chlorination was performed on the raw wastewater and on the reverse osmosis brine. The raw wastewater consisted of blast furnace scrubber blowdown that had been clarified and passed through a cooling tower. The reverse osmosis brine consisted of the byproduct from chemically treated, dual media filtered raw wastewater that had been passed through a reverse osmosis unit.
4. Mixing: Rapid mixers were placed in each of the reaction chambers to insure adequate contact between the wastewater and the chemical applied. Following rapid mix, the waste was passed through a clarifier to remove suspended solids.
5. Chlorine Application: From the bench studies, ammonia was found to be the limiting parameter for meeting BAT limitations. A chlorine to ammonia ratio of 8.7:1 was found necessary to reduce

Date						
Run No.						
Start time						
End time						
Flow rate, gpm						
Influent Data						
pH						
Temp., °C						
NH ₃ conc., mg/l						
Cl ₂ residual, mg/l						
Tank number						
NaOCl						
Feed conc., %						
Addition rate, ml/min						
Cl ₂ applied, mg/l						
NaOH						
Feed conc., %						
Initial vol., gal.						
Final vol., gal.						
Total used, gal.						
pH adjusted to						
Tank number						
HCl						
Feed conc., %						
Initial vol., gal.						
Final vol., gal.						
Total used, gal.						
FeCl ₃						
Feed conc., %						
Addition rate, ml/min						
Tank number						
pH						
Cl ₂ residual, mg/l						
NH ₃ residual, mg/l						
Clarifier Effluent						
pH						
Cl ₂ residual, mg/l						
NH ₃ residual, mg/l						
Temp., °C						

Figure 12. Alkaline chlorination data sheet.

ammonia concentrations to acceptable limits. During the pilot testing, the chlorine to ammonia ratio was varied during each run so that the optimum bench study ratio was bracketed.

Pilot Test Procedure and Data Acquisition--

The alkaline chlorination process was run continuously while testing the raw wastewater and on a batch basis for treating of the reverse osmosis brine. Before each test, a chlorine to ammonia ratio was selected. The testing interval was called a run and assigned a number. Run lengths were generally from 8 to 24 hours. There were occasional runs of longer or shorter duration.

CHEMICAL TREATMENT

Introduction

The mobile research trailers used in this project were equipped with two chemical clarification-sedimentation systems. A 1.8 meter (6.0 ft) diameter clarifier was used to remove suspended solids and turbidity to prepare the wastewater for downstream processes, and a 1.3 meter (4.25 ft) diameter clarifier was utilized as a polishing step to reduce suspended solids and turbidity from upstream processes. Chemical addition to the 1.8 meter (6.0 ft) diameter pretreatment clarifier typically consisted of metal salts and polymer, while the 1.3 meter (4.25 ft) diameter polishing clarifier used polymer to enhance settling. A schematic diagram of the treatment trains using chemical clarification has been shown previously in Figure 4.

Description of the Pilot Equipment

Physical characteristics of the 1.8 meter (6.0 ft) diameter pretreatment clarifier are shown below:

Overall Diameter	1.83 m	(6.0 ft)
Net Surface Area For Clarification	2.63 m ²	(28.3 ft ²)
Clarifier Volume	2.73 m ³	(96.3 ft ³)
Flocculation Volume	0.23 m ³	(8.0 ft ³)
Rapid Mix Volume	0.045 m ³	(1.6 ft ³)
Clarifier Detention Time	1 hr @ 0.045 m ³ /min (12 gpm)	
Rapid Mix Detention Time	1 minute @ 0.045 m ³ /min (12 gpm)	
Flocculator "G"'s	0-180 m/sec/m (0-180 ft/sec/ft) max. variable speed turbine	

The flocculation chamber was a cylinder located in the center of the clarifier. Mounted in the center of this chamber was a 30.5 centimeter (12 inch) diameter turbine flocculator. The blades of the turbine were mounted to produce axial flow. Feed flow was piped into the draft tube and discharged above the turbine. The flocculated wastewater proceeded down the draft tube and into the settling zone. The sludge was removed automatically by a

timer controlled pump while the clarified effluent was discharged over a weir. Six sampling points were located every 15.9 centimeters (6.25 inches) up the side of the clarifier. These taps are used to determine sludge depth.

The physical characteristics of the 1.3 meter (4.25 ft) diameter polishing clarifier are outlined below.

Overall Diameter	1.3 meter	(4.25 ft)
Net Surface Area For Clarification	1.32 m ²	(14.2 ft ²)
Clarifier Volume	1.53 m ³	(47.7 ft ³)
Clarifier Detention Time	1.0 hr @ 0.0225 m ³ /min (6 gpm)	

Flow to the unit entered the center-well and was distributed radially with sludge being removed manually from the bottom. The effluent flowed out over 90° V-notch weirs. There was no flocculation chamber in the polishing clarifier.

Chemical Feed System

Chemicals were introduced into the clarifiers' feed water either in-line at tees or into the rapid mix tanks directly. Betz 1100 polymer, ferric chloride, and alum were typically used during clarification.

The chemical feed pumps were operated in the "auto" mode or "hand" mode. In the auto mode the chemical pump operated in synchronization with the feed pump. The hand mode allowed the pump to operate independent of the feed pump. The pH control system maintained a steady pH into the clarifiers. High and low pH alarms were also provided. Chemical feed flow rates were periodically checked in the field by the operator.

Experimental Program

The objective of the chemical clarification step was to produce an effluent low in suspended solids and turbidity. Based on the results of the bench testing program, alum and ferric chloride were found to be effective coagulants. A low charge density anionic polymer, Betz 1100, served as an effective coagulant aid.

Wastewater loadings to the two clarifiers were generally dictated by hydraulic requirements of upstream or downstream processes, depending on the treatment train being investigated.

The pretreatment clarifier was used to (1) settle alkaline chlorination effluent prior to dechlorination by activated carbon; and (2) pretreat the raw wastewater prior to filtration and reverse osmosis. The sludge blanket from the unit was maintained at a constant level by using a sludge pump connected to an automatic timer.

The polishing clarifier was utilized to (1) settle residual solids present in the ozonated effluent; (2) settle ozonated reverse osmosis brine solids; and (3) pretreat the alkaline chlorinated reverse osmosis brine prior to activated carbon treatment. Sludge depth was typically controlled by manual wasting.

For the majority of the test program, the clarification systems were operated continuously (24 hours per day, seven days per week). During the day, personnel were available to monitor the performance of the unit, whereas during the evenings the unit ran unattended. A typical data sheet used during the test program is presented in Figure 13. Photographs of the chemical treatment system have been shown previously in Figures 9 and 10.

DUAL MEDIA FILTRATION

Introduction

Filtration is widely utilized for wastewater treatment both in upgrading existing conventional plants and in designs of new advanced treatment facilities. It is one of the most widely used processes for separation of wastewater solids.

In application, filtration may serve both as an intermediate process to prepare wastewater for further treatment (such as carbon adsorption, or reverse osmosis) and as a final polishing step following other processes.

Granular media filtration involves passage of water through a bed of granular material with resulting deposition of solids. Eventually the pressure drop across the bed becomes excessive or the ability of the bed to remove suspended solids is impaired. Cleaning is then necessary to restore operating head and effluent quality to acceptable levels. Most filters operate on a batch basis, the entire unit being removed from service for periodic cleaning.

The time in service between cleanings is termed the run length. The head-loss at which filtration is interrupted for cleaning is called the terminal headloss.

Filter design involves selection of the following filter characteristics:

1. Filter configuration.
2. Media sizes, depths and materials.
3. Filtration rate, l/min/m^2 (gpm/ft^2).
4. Terminal headloss, m of water (ft of water).
5. Method of flow control.
6. Backwashing design features.

The major goal in design is to achieve effluent quality objectives at low capital and operating costs. The most important characteristic in determining capital costs is the filtration rate, which fixes the filter size.

CHEMICAL TREATMENT - DATA SHEET #1

Date						
Time						
Feed Source						
Raw Flow - Clarifier 1						
pH						
Turbidity						
Temperature						
gpm						
Chemical Addition						
FeCl ₃ - ml/min						
% stroke						
ppm						
Polymer - ml/min						
% stroke						
ppm						
Effluent Flow						
Flow totalizer						
Time						
gpm						
pH						
Turbidity						
Temperature						
Clarifier - 1						
Turbine rpm						
Control setting						
Sludge wasting						
Time						
Rate						
Blanket depth						
Settled volume/time						
Sample ESD #						
Clarifier influent						
Clarifier effluent						
Sludge						

Figure 13. Chemical treatment data sheet.

Operating costs are affected primarily by filtration rate, terminal headloss, media characteristics and backwash design. The first three filter characteristics determine the cost of power for operating head and the production of the filter per run. The backwash design determines the cost per cleaning of operator attention, washwater pumping, air scouring (compressor operation) and treatment of dirty washwater. The cost of cleaning per unit volume treated (cost per cleaning divided by production per run) depends on all four factors.

Wherever possible, designs should be based on pilot filtration studies of the actual waste. Such studies are the only way to assure:

1. Meaningful cost comparisons between different filter designs capable of equivalent performance, i.e., producing the same output quantity and quality over the same time period.
2. Most economical selection of filter rate, terminal headloss and run length for a given media application.
3. Definite effluent quality performance for a given media application.

Pilot studies are also used for determining effects of pretreatment variations or for characterizing filterability in terms of performance attainable with a specific filter design.

The wastewater tested was raw blast furnace blowdown and chemically clarified blowdown. During the second phase of testing, chemical clarification-sedimentation was necessary to aid in the removal of residual suspended solids and turbidity for downstream processes.

Description of Pilot Equipment

An automated, downflow, granular media filter with an effective area of 0.09 m^2 (1.0 ft^2) was used in this study. The filter compartment internal diameter is 0.35 meters (13.6 in.) and the height is 2.1 meters (7 ft). The vessel is rated for 35,155 kg/m (50 psi) operation, and is assembled in three flanged sections. The bottom section contains a filtrate compartment, separated from the filter media compartment by a false bottom. Connections for filtered product water, backwash water and air are located in the filtrate compartment. Three plastic underdrain strainers are located in the false bottom. Each strainer head measures 6.3 cm (2.5 in.) in diameter with 0.025 cm (.010 in.) slots. The strainer also had a 5.2 cm (6 in.) long air metering tube extending into the filtrate compartment. Holes in the air metering tubes provide for uniform distribution of air to each strainer during air backwash. The strainers are designed for use in the sand media without a gravel support.

Above the filtrate compartment the filter media compartment consisted of a flanged section and a flanged cover. A flanged access port is located in the lower portion of the section. Screened pressure taps are provided at 10.2 cm (4 in.) intervals, starting 7.6 cm (3 in.) from the base plate. The top cover contains connections for the feed and backwash overflow.

Filtrate was collected in a 1,900 l (500 gal.) tank to provide a continuous supply to downstream processes during filter backwash periods.

Flow to the filter was pumped by a 84.9 BTU/min. (2 HP), enclosed impeller, centrifugal pump. Backwash water was pumped from the 1,900 l (500 gal.) tank by a 31.8 BTU/min. (3/4 HP), enclosed impeller, centrifugal pump and an air compressor was provided for backwash air scour.

Constant flow rate could be maintained with a flow control valve on the filtered effluent. Backwash could be initiated by excessive headloss, timed interval or manually. In actual practice, only the manual mode was utilized so that final pressure readings could be taken before backwashing. Once backwash is initiated, the sequence proceeds automatically with draw-down of the liquid above the filter bed, air scour, water wash, and a short idle interval to allow the media to settle. The filter is then returned to normal service. Valves for automatic functions are motor actuated.

Other instrumentation includes a filtrate flow rate recorder and a backwash flow meter. Headloss within the filter media was monitored at the pressure taps in the side of the filter vessel. Each pressure tap was valved and connected through a manifold to an array of three differential pressure meters. The ranges of these instruments were 2.54-63.5 cm (1-25 in.) water, 5-254 cm (2-100 in.) water and 0-42,186 kg/m² (0-60 psi). Headlosses within the media were always based on the pressure above the media. The pilot filter before installation in the research trailer is shown in Figure 14.

Experimental Program

The objectives of pilot filter tests were to obtain sufficient information for full-scale design. Of vital interest are performance characteristics for determining effluent quality and design parameters to insure the most cost-effective equipment size.

Performance characteristics include effluent quality and the consistency of effluent during varying feed conditions. The filter surface area required to process the design flow, media selection, backwash water processing requirements, feed preconditioning requirement, power requirements, and filter cleaning requirements are examples of design parameters which can be obtained from pilot testing.

Pertinent variables include the following:

1. Filtration rate, l/min/m² (gpm/ft²), filter area.
2. Feed characteristics.
 - a. suspended solids, mg/l
 - b. size distribution
 - c. resistance to shear
3. Media selection.
 - a. composition
 - b. size distribution

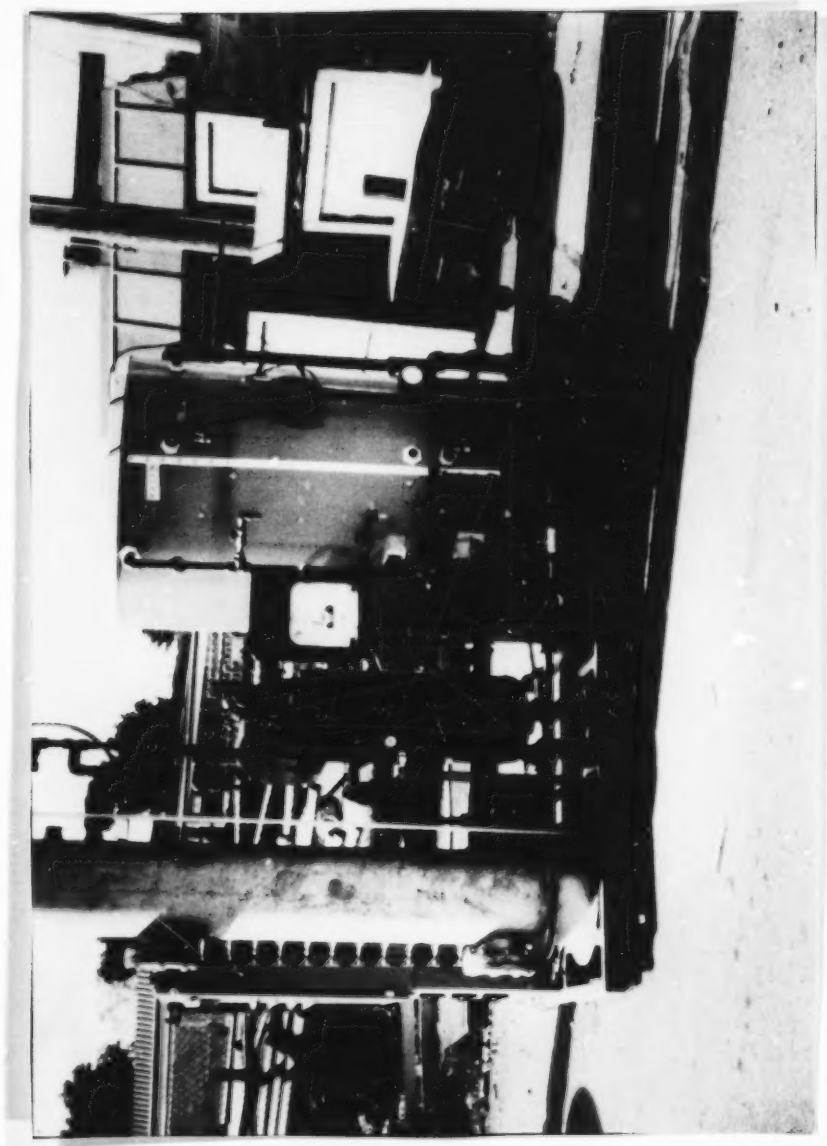


Figure 14. Dual media pilot filter.

- c. depth
- 4. Backwash.
 - a. rate
 - b. duration
 - c. procedure
- 5. Mode of filter operation.
 - a. constant head - declining rate
 - b. constant head - constant rate
 - c. variable head - constant rate
 - d. variable declining rate
- 6. Terminal headloss.
- 7. Chemical preconditioning.

Pilot testing enables the researcher to define how these independent variables affect the following dependent variables.

- 1. Measurement of production.
 - a. weight of solids captured per unit filter area as a function of headloss.
 - b. net volume of water filtered as a function of headloss.
- 2. Filtrate quality.
- 3. Cause of filtration termination.
 - a. headloss
 - b. solids breakthrough
- 4. Depth of solids penetration as indicated by a headloss profile of the filter bed.
- 5. Effectiveness of backwash.

The data obtained during each filter run were recorded on a standard data sheet. Routine recordings included time, flow rate, headloss profiles, turbidity, pH, and temperature (in and out). A sample data sheet is shown in Figure 15.

Test Conditions

Based on prior lab bench tests and accepted filtration practice, the following test conditions were selected.

- 1. Filtration Rate: The filtration rate was varied from 122 to 407.4 $\ell/\text{min}/\text{m}^2$ (3-10 gpm/ft^2) throughout the study period. For each type of feed the test conditions generally included filtration rates of 122, 204, 326 and 407.4 $\ell/\text{min}/\text{m}^2$ (3, 5, 8, and 10 gpm/ft^2).
- 2. Feed Characteristics: This variable is determined by the process preceding the filters. During the study, the dual media filter processed blast furnace cooling tower blowdown and chemically clarified blowdown. During the first treatment train, the blowdown was filtered directly to determine suspended solids removal without chemical addition.

Waste Stream: _____ Hydraulic Loading: _____
 Plant Operation: _____ Operator: _____
 Run Number: _____ Remarks: _____

[illegible][illegible]

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During the reverse osmosis study, the blowdown was clarified with alum, prior to filtration. At that time, alum was added to the blowdown, flocculated and settled in the pilot clarifier. Polymer was also added.

3. **Filter Media:** Bench scale testing early in the program indicated the best effluent quality could be achieved with dual media filtration using anthracite with an effective size of 1.0 millimeter over sand with an effective size 0.5 millimeter. For the pilot program, anthracite coal was purchased from The Reading Anthracite Coal Company, Pottsville, Pennsylvania. The purchased specification was for a material with 0.80 to 1.20 millimeter effective size and maximum uniformity coefficient of 1.65.

Filter sand was obtained from the Red Flint Sand Company, Eau Claire, Wisconsin. The purchased specification was for 0.45 to 0.55 millimeter effective size and a uniformity coefficient of 1.4. Before loading the media, representative samples were taken by the "cone and quartering" method for sieve analysis. The results are presented in Figure 16. The anthracite media used in the pilot program had an effective size of 0.94 millimeter with a 1.60 uniformity coefficient, while the sand media effective size was 0.47 millimeter with 1.34 uniformity coefficient.

On loading the media, a known quantity of sand was placed in the filter and backwashed at 611 l/min/m^2 (15 gpm/ft^2) for 30 minutes. To remove fine material, the top 5 cm (2 in.) was removed. Next, a known amount of anthracite was placed on the sand and backwashed and 5 cm (2 in.) removed. The total media depth was then recorded.

4. **Backwash:** A relatively mild degree of backwashing was used throughout the study. The procedure was to first draw the water level down 30.5 cm (12 in.) from the overflow trough. Next, backwashing was initiated (203.7 l/min/m^2) [5 gpm/ft^2] while simultaneously agitating with air at $1.5 \text{ standard m}^3/\text{min/m}^2$ (5 scfm/ft^2) for two minutes. After turning off the air, filtered effluent was used to backwash the filter at 611 l/min/m^2 (15 gpm/ft^2) for eight minutes. Each backwash was initiated manually so that a given filter run would correspond to a sampling interval.
5. **Mode of Filter Operation:** All tests were run under pressure with constant head and constant rate.
6. **Terminal Headloss:** The pilot filter was operated at a pressure of about $35,156 \text{ kg/m}^2$ (50 psig). Data taken during each run allowed evaluation of both pressure and gravity filtration.

Pilot Test Procedure and Data Acquisition

The filter was operated continuously throughout the study. Each interval between backwashings was called a run and assigned a number. Run lengths

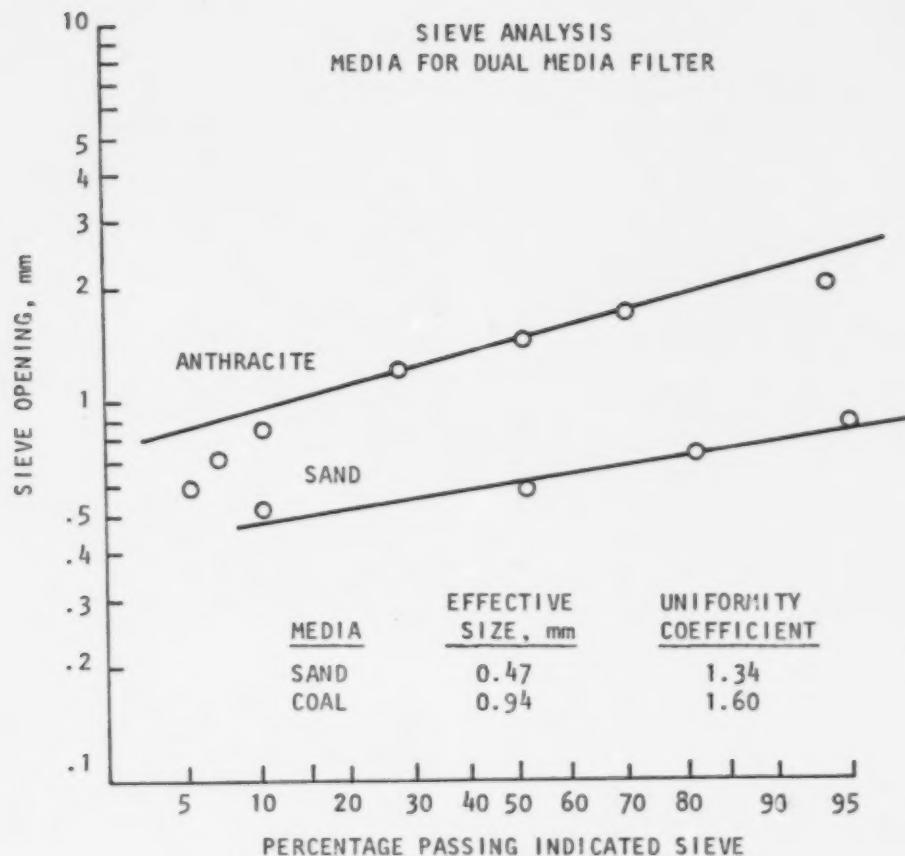


Figure 16. Filter media sieve analysis.

were generally determined by the sampling schedule and were approximately 24 hours long. There were occasional runs of longer or shorter duration as warranted by the headloss through the media. At the beginning of each run the flow controller was set to the desired rate and initial headloss readings taken. Periodically during the run, overall headloss was recorded and turbidity of the feed and filtrate was measured. Headloss profiles through the media were recorded at intervals through the run and at termination.

Samples of feed and filtrate were composited for each run for suspended solids determination. At the termination of a run the filter was backwashed and prepared for the next run. Occasionally, samples of the backwash water were taken at intervals during the water wash cycle and analyzed for suspended solids.

MAGNETIC FILTRATION

Introduction

The principal of magnetic separation dates back many years to application of removing tramp iron from nonferrous products and separating or concentrating iron-bearing ores. In recent years, this technology has been applied in "wet applications" of treating wastewater.

The greatest success has been achieved with wastewater with large, highly magnetic particles. However, moderate success has also been obtained with weakly magnetic submicron sized particles under high gradient magnetic fields. The use of a seeding material such as magnetite with a flocculating agent has proven helpful in these more difficult applications.

Magnetic filtration involves the passage of water through a magnetically charged matrix in an inner separating chamber. The matrix is magnetically charged by an electromagnet contained in an outer sealed chamber. If the magnetic attraction of the particles for the matrix is greater than viscous drag of flowing liquid, the particles will deposit on the matrix. Some particles also may become mechanically trapped. Increasing deposition will result in increased pressure drop across the media and impaired particle removal. At this point, the matrix can be cleaned by de-energizing the electromagnet and backwashing the matrix.

Most filters operate on a batch basis, the entire unit being removed from service for periodic cleaning.

The time in service between cleanings is termed the run length. The headloss at which filtration is interrupted for cleaning is called the terminal headloss.

Filter design involves selection of the following filter characteristics:

1. Filter configuration.
2. Media sizes, depths and materials.
3. Filtration rate, l/min/m^2 (gpm/ft^2).
4. Terminal headloss, m of water (ft of water).
5. Method of flow control.
6. Backwashing design features.

The major goal in design is to achieve effluent quality objectives at low capital and operating costs. The most important characteristic in determining capital costs is the filtration rate, which fixes the filter size. Operating costs are affected primarily by filtration rate, terminal headloss, media characteristics and backwash design. The first three filter characteristics determine the cost of power for operating head and the production of the filter per run. The backwash design determines the cost per cleaning of operator attention, wash water pumping, the treatment of dirty wash water. The cost of cleaning per unit volume treated (cost per

cleaning divided by production per run) depends on all four factors.

Wherever possible, designs should be based on pilot filtration studies of the actual waste. Such studies are the only way to assure:

1. Meaningful cost comparisons between different filter designs capable of equivalent performance, i.e., producing the same output quantity and quality over the same time period.
2. Most economical selection of filter rate, terminal headloss and run length for a given media application.
3. Definite effluent quality performance for a given media application.

Pilot studies are also useful for determining effects of pretreatment variations or for characterizing filterability in terms of performance attainable with a specific filter design.

Description of Pilot Equipment

During the testing, the steel wool matrix was chosen instead of the grid matrix because of its greater surface area per unit volume. The steel wool used had a cross sectional area of 71 cm^2 (11 sq in.).

The electromagnet of the filter generated a magnetic field with a flux density of 1,000 gauss. At this operating level, it used 5 amps of 115V alternating current.

Filtrate from the unit was collected in a 1,900 l (500 gal.) tank to provide a continuous supply to downstream processes during filter backwash periods.

Flow to the filter was pumped by a 84.9 BTU/min. (2 HP), enclosed impeller, centrifugal pump. Backwash water was pumped from the 1,900 l (500 gal.) tank by a 31.8 BTU/min. (3/4 HP), enclosed impeller, centrifugal pump.

Constant flow rate to the unit was maintained with a flow control valve on the filtered effluent.

Backwash could be initiated by excessive headloss, timed interval or manually. In actual practice, only the manual mode was utilized so that final pressure readings could be read before backwashing.

Once backwash is initiated, the sequence proceeds automatically with the feed pump stopping, coil power being interrupted, appropriate valves opening or closing, backwash pump starting. After backwashing, the filter returns to normal service. Other instrumentation includes a filtrate flow rate recorder and a backwash flow meter. Headloss within the filter matrix was monitored with pressure taps at the influent and effluent points of the filter.

Test Conditions

The filter was tested on raw blast furnace wastewater, chemically pretreated blast furnace wastewater, and "magnetite seeded" chemically pretreated blast furnace wastewater. During the program, the flow through the filter was varied from a high of 2,648 l/min/m^2 (65 gpm/ft^2) to a low of 534 l/min/m^2 (13 gpm/ft^2). The manufacturer's rated capacity of the unit was 2,648 l/min/m^2 (65 gpm/ft^2) equivalent to 18.9 l/min (5 gpm).

The filter was operated at selected intervals, throughout the study. Each interval between backwashing was called a run and assigned a number. Run lengths were generally determined by effluent quality or headloss through the matrix. At the beginning of each run the flow controller was set to the desired rate and initial headloss readings taken. Periodically during the run, overall headloss was recorded and turbidity of the feed and filtrate was measured. Headloss profiles through the media were recorded at intervals through the run and at termination.

Samples of feed and filtrate were composited for each run for suspended solids determination. At the termination of a run the filter was backwashed and prepared for the next run. Occasionally samples of the backwash water were taken at intervals during the water wash cycle and analyzed for suspended solids.

During the run, intermittent grab samples of the feed and filtrate were taken for field analysis of turbidity, pH and temperature. The data were recorded on a standard data sheet as shown in Figure 17.

REVERSE OSMOSIS

Introduction

Reverse osmosis (RO) was one of the blast furnace blowdown treatment technologies selected to investigate upgrading the wastewater to meet BAT discharge limitations. Reverse osmosis is an advanced waste treatment process capable of removing a significant portion of suspended, colloidal and dissolved contaminants present in wastewater. Contaminant removal is obtained by the filtration of the waters through a semi-permeable membrane. Typical pore sizes of the membrane range from 20-50 angstrom units, while pressures of 281,240 - 421,860 kg/m^2 (400-600 psig) are usually encountered. The resultant treated, or product water, as it is referred to, is typically low in dissolved organic and inorganic contaminants.

Pretreatment of the waste prior to reverse osmosis application was required. Chemical flocculation and sedimentation, using alum and polyelectrolyte followed by dual media filtration were typically required to meet the feed quality standards.

Waste Stream: Raw Wastewater Feed _____

Plant Operation: _____

Run Number: _____

Hydraulic Loading: _____

Operator: _____

Remarks: _____

[illegible]

Figure 17. Magnetic filtration data sheet.

Description of Pilot Equipment

There are three types of commercially available RO membranes; the tubular, spiral wound, and hollow fiber configurations. Only a spiral wound unit was evaluated during the study because the bench testing phase of the program indicated that the pretreatment available was not adequate for successful operation of a hollow fiber unit. Consideration was not given to the tubular configuration due to low membrane surface to unit volume ratio and high operating expense.

In the spiral type design, a flat sheet of membrane is used to cover each side of a sheet of porous water-conducting material. The membrane is sealed on the two long edges and one of the other ends to form an envelope. The remaining end is attached to a perforated tube to collect the product water. A spacer is placed on the membrane, then the whole assembly is rolled up around the center collection tube. Feed water enters the module and flows through the channels produced by the spacer. A portion of the feed permeates the membrane and flows radially through the porous material to the collection tube. See Figure 18 for more detail.

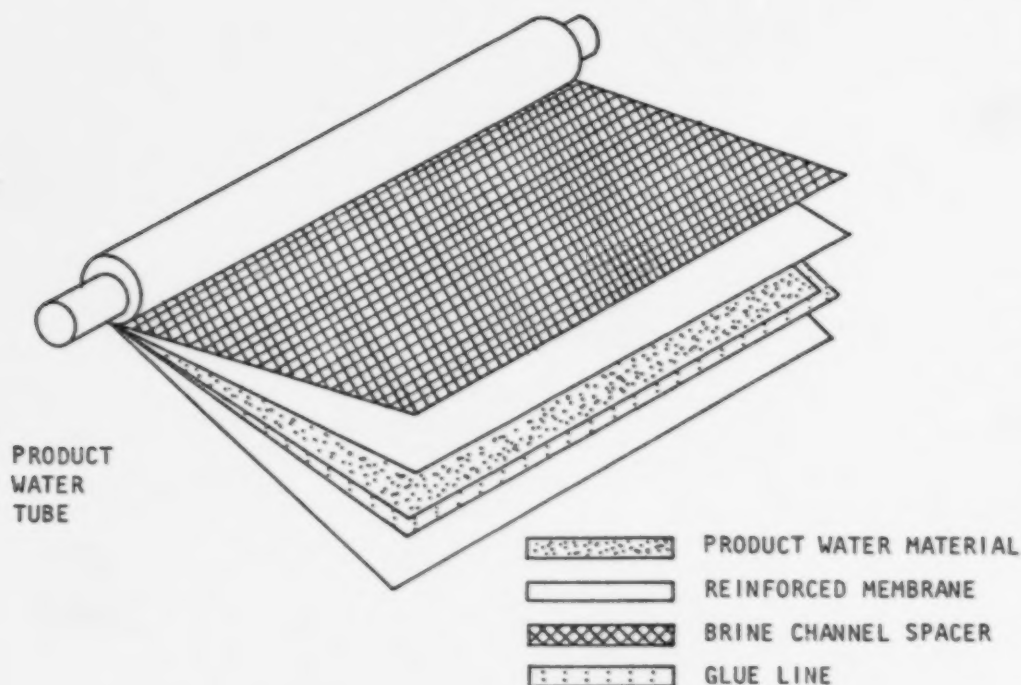


Figure 18. Interior details of a spiral wound membrane.

The spiral wound reverse osmosis membranes used in the study were thin-film composite (TFC) 4600 polyamide (PA) membranes purchased from Fluid Systems Division, UOP Inc. The TFC 4600 PA membranes manufactured by Fluid Systems Division recently became commercially available and represent a new advance in reverse osmosis membrane technology. TFC membranes consist of an ultrathin film of a semipermeable polymer formed upon the finely porous surface of a highly water-permeable support membrane. See Figure 19 for greater detail. Independent selection and development of the thin film and the porous supporting membrane allows each component to be optimized for its specific function.

Reverse osmosis systems utilizing cellulose acetate membranes have restricted areas of application because of limitations associated with these membranes. These limitations include low water permeability, poor chemical stability, temperature limitations, and associated flux decline due to compaction. Polyamide membranes have several advantages over cellulose acetate membranes. Their greater water permeability permits higher product water production with comparable or better quality, PA membranes have greater chemical stability allowing an operating pH range of 2-12 and the use of strong cleaning agents for more effective membrane cleaning. They exhibit higher inorganic and organic contaminant rejection. PA membranes have demonstrated the capability to operate at elevated temperatures, 55 °C (131 °F), without significant flux decline due to compaction. The RO system used during the pilot program is shown in Figure 20. The unit consists essentially of a booster pump, the high pressure pump, the module containing three spiral wound membranes in series, a mixing tank for pH adjustment, and the necessary piping and controls to operate and monitor the system. The unit is equipped with a brine recycle loop. A 10 micron cartridge filter was provided upstream of the high pressure pump to protect the pump and membranes from any large suspended material present in the feed water. A photograph of the RO pilot unit is shown in Figure 21.

Experimental Program

The three main objectives of the experimental program were to determine:

1. The quality of the product water with respect to BAT limits.
2. The rate of fouling of the membranes.
3. The effectiveness of various cleaning agents in restoring the fouled membranes to their original condition.

Because of time limitations, it was not possible to study other design parameters such as additional pretreatment, different types of membranes, etc. The pretreatment and the operating conditions for the system were based either on manufacturer's recommendations or results of other pilot or full-scale programs.

The reverse osmosis test program was designed to evaluate the performance of the RO system under two sets of operating conditions. This dual phase study accurately predicted the performance of a full size RO system. The first phase provided an "optimistic" prediction of the product water quality

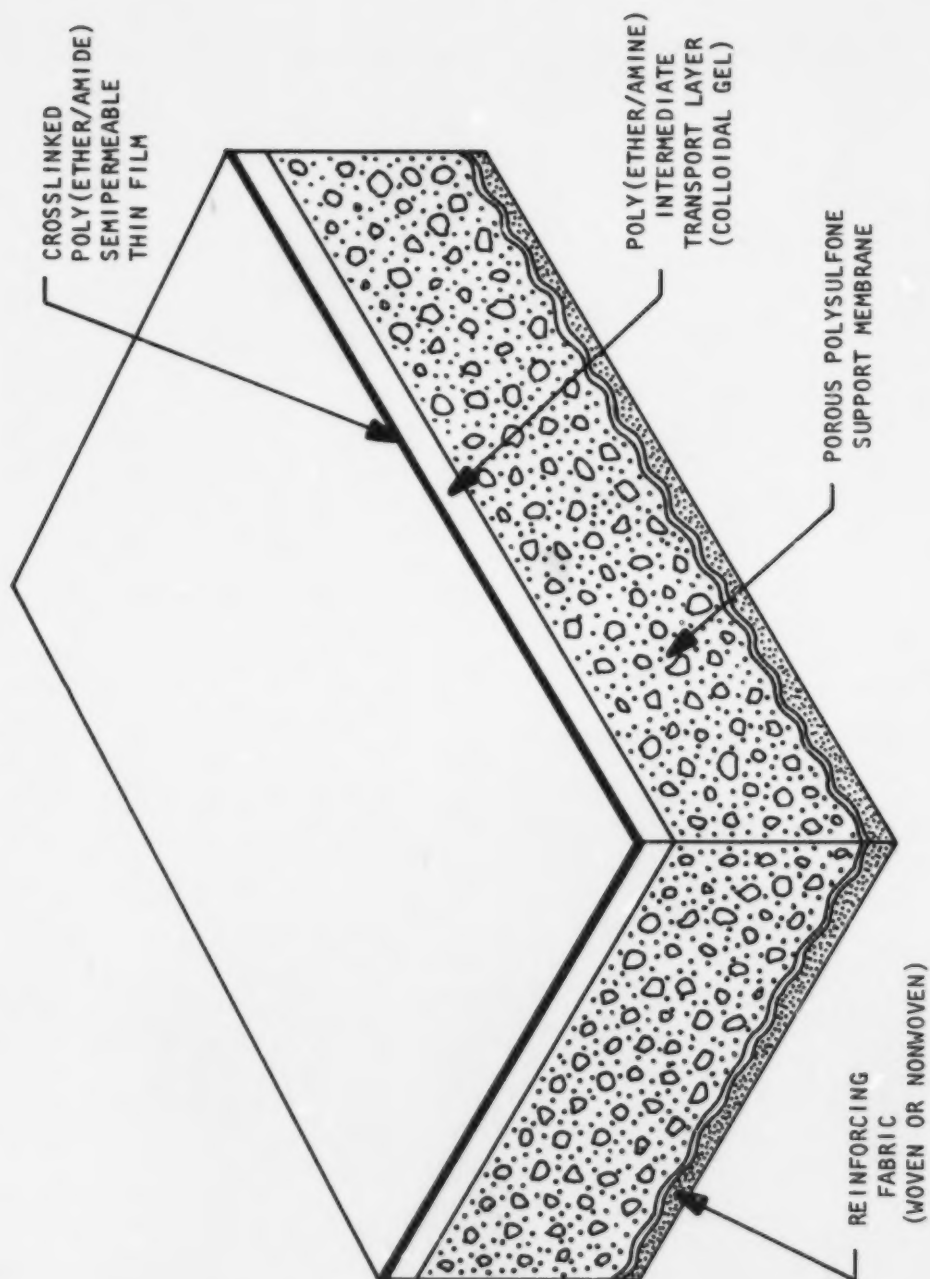


Figure 19. Cross sectional drawing of a polyamide thin-film composite membrane.

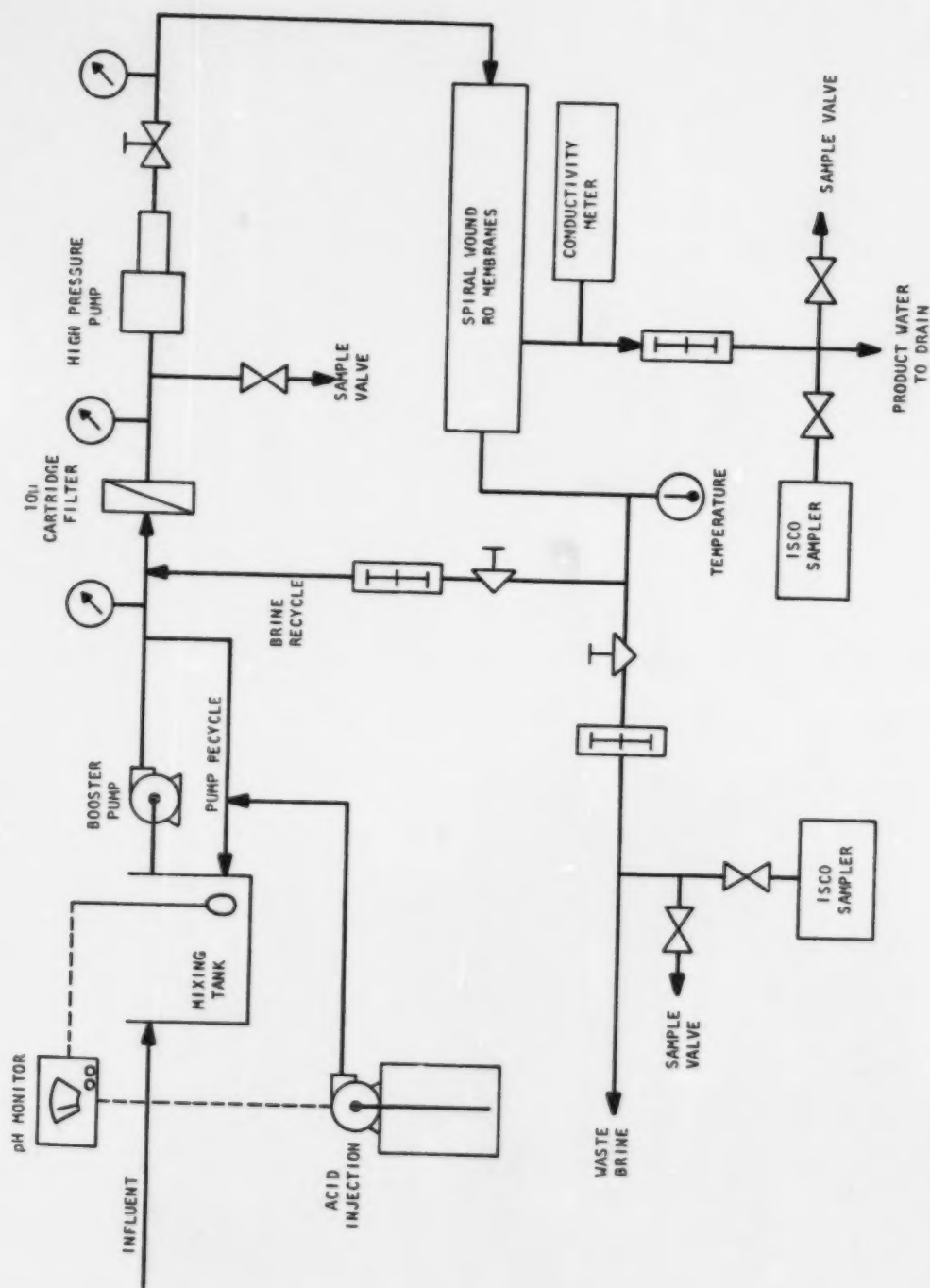


Figure 20. Schematic of pilot reverse osmosis unit.



Figure 21. Reverse osmosis pilot system.

that could be expected from a full size unit. The product water produced was similar in quality to that produced in the first series of membranes in a full size unit (representing 25 percent of the total volume of product water). Figure 22 shows a schematic of a full size RO system. The membranes were operated at product water recoveries of 25-30 percent in the first phase.

The second phase provided a "worst case" prediction of product water quality. The product water quality would be similar to that produced in the last series of membranes in a full size unit (representing 17 percent of the total volume of product water). The fouling frequency and the effectiveness of the cleaning solutions were determined in this stage of the program. The membranes were operated at product water recoveries of 70-75 percent by using the brine recycle loop provided with the unit.

During the test program, the RO unit was run 24 hours a day, 7 days a week unless mechanical or operational problems forced a shutdown. Manpower was available during the day to operate the units. The RO unit ran unattended over night. Operating data, such as temperature, pressure, flow rate, conductivity, and pH of the feed, brine and product flow streams, were recorded several times daily. Samples for laboratory analysis consisted of run composites and grab samples. A typical data sheet used to record operational parameters is presented in Figure 23.

The operation of a spiral wound unit consists of adjusting the feed pressure to obtain and maintain the rated product flow of the unit. On initial startup, the feed pressure is adjusted to obtain the desired product flow rate. As the unit is operated and the membranes foul, the pressure is increased to maintain the rated output of the membranes. When the feed pressure has increased by 15-20 percent, the membranes are considered to be fouled and cleaning is necessary.

Two cleaning solutions were typically evaluated during the study. For removing organic fouling material a 1.5 percent soda ash solution was used. When the fouling material was inorganic in nature, the membranes were cleaned with a 2 percent citric acid solution. The citric acid and soda ash solutions are recommended for cleaning TFC membranes (3). Figure 24 presents a typical data sheet used to identify the cleaning solution used and the change in operational parameters due to cleaning.

OZONATION

Introduction

Ozone, the peroxidic molecule O_3 , is a powerful oxidant which can oxidize most of the contaminants in wastewater. Ozone has been used for disinfecting and purifying drinking water in other countries for many years (4), and has attracted considerable interest in the United States over the past two decades, particularly for waste treatment applications. The attractiveness of ozone for waste treatment arises from certain advantages in its

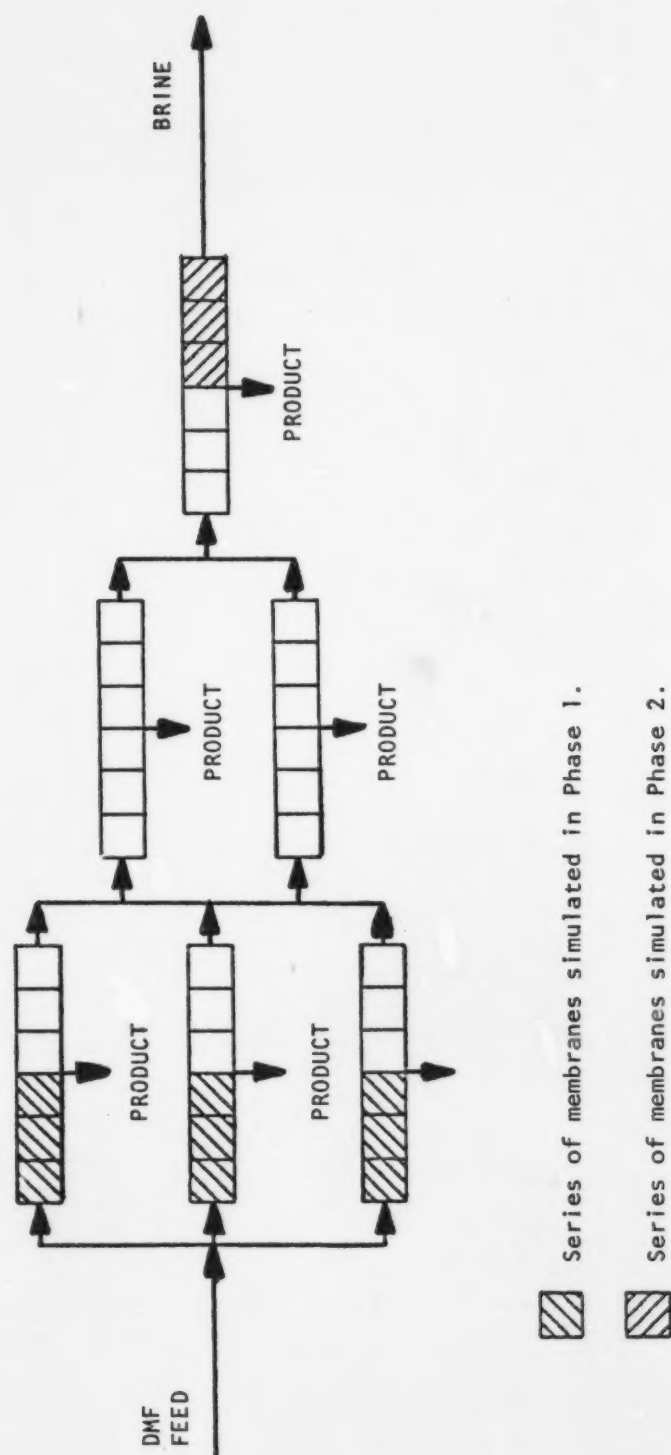


Figure 22. Schematic of staged RO system.

Date					
Time					
Cumulative Hours					
Temperature - °F					

Phase / Run					
Feed Turbidity					
Cartridge Filter Size (μ)					
Cartridge Pressure Drop					

Pressure - psi	B	A	B	A	B	A	B	A	B	A
Feed										
Reject										
Initial Feed Pressure										

Flow - gpm	B	A	B	A	B	A	B	A	B	A
Feed										
Brine										
Product										
Recycle										
% Recovery										

Conductivity					
Feed					
Brine					
Product					
Salt passage % (P/Fx100)					

pH					
Feed					
Brine					
Product					
Acid carboy weight ^(initial) / _{final}	/	/	/	/	/
Operator					

Figure 23. Reverse osmosis operating data sheet.

REVERSE OSMOSIS CLEANING DATA

Clean After Run _____ Phase _____
 Date of Cleaning: _____
 Time of Cleaning*: _____
 Cleaning Done By: _____

Cleaning Solution

Gallons Prepared: _____
 Cleaner Used: _____
 Amount of Cleaner: _____
 pH Adjusted With: _____
 Amount: _____
 pH of Cleaning Solution: _____

Cleaning Summary

	Before	After	Change	
Temp. °F				Flushed for _____ hrs
Feed Pres.-psi				Flushed at _____ gpm
Reject Pres.-psi				Solution temp. _____ °F
ΔP - psi				Comments:
Feed Flow-gpm				
Brine Flow-gpm				
Product Flow-gpm				
Recycle Flow-gpm				
Recovery %				
Cond. Feed				
Cond. Brine				
Cond. Product				
Salt Passage				

* Accumulative hours from control panel - not time of day.

Figure 24. Reverse osmosis cleaning data sheet.

application (5):

1. Many wastewater components that are toxic, carcinogenic, or taste and odor producing compounds can be oxidized by ozone.
2. Except for oxygen, no residual chemical additives remain in the product water.
3. Ozonation produces a high dissolved oxygen concentration (DO) in the product water.

Pilot Plant Description

The ozonation studies were conducted using a pilot unit manufactured by the Ozone Technology Group, Emery Industries, Inc., Cincinnati, Ohio. A schematic diagram of the unit is presented in Figure 25. Photographs of the system are shown in Figure 26.

Wastewater was pumped to a mixing tank where sodium hydroxide (NaOH) was added to raise the pH to the 11.0-12.0 range using a pH controller. Following pH adjustment, the wastewater was pumped through three ozone contactors in series. The wastewater flow is counter-current with the ozone rich gas for more efficient diffusion of the gas into the liquid. The three reinforced fiberglass contactors are 0.5 m (20 inches) in diameter, 2.4 m (8 ft) high, and have a water depth of 1.5 m (5 ft). The contactors contain a total liquid volume of 1,135 l (300 gal.) for a contact time of 60 minutes at 18.9 l/min (5 gpm). The liquid level in the contactors is controlled by a standpipe on the effluent piping of contactor No. 3. Sample taps are provided on the piping between contactors No. 1 and 2 and 2 and 3. Effluent from the third contactor flowed by gravity to a mixing chamber where hydrochloric acid (HCl) was added to reduce the pH to 7.0. After neutralization, the wastewater was flocculated with Betz 1100 polyelectrolyte and clarified.

The ozone gas generating system uses air as feed. A compressor, rated at 12,700 standard cc/sec (27 scfm) at 703,000 kg/m² (100 psi), and a water cooled aftercooler provide a cool, compressed air to a desiccant air dryer. The air dryer produced -51 °C (-60 °F) dewpoint air which was fed to the ozone generator at 10,500 kg/m² (15 psi). The ozone generator was a water cooled Emerzone Model 9275, containing 31 dielectrics, capable of producing 12.2 kg/day (27 lb/day) of ozone. The ozone rich gas was sent to the contactors by a parallel piping system. The ozone concentration of the gas can be measured using a sample tap on the gas piping. Each contactor has a separate throttling valve and gas flow indicator to control and measure the gas flow. Fine bubble diffusers are used to diffuse the ozone rich gas into the wastewater. A sample tap is provided on the off-gas piping of each contactor allowing the measurement of residual ozone present in the off-gas. The off-gas from the three contactors is sent to an entrainment separator to remove entrained moisture. Finally an ozone destruction unit operating at 315 °C (600 °F) destroys the residual ozone in the off-gas stream before discharging the gas to the atmosphere.

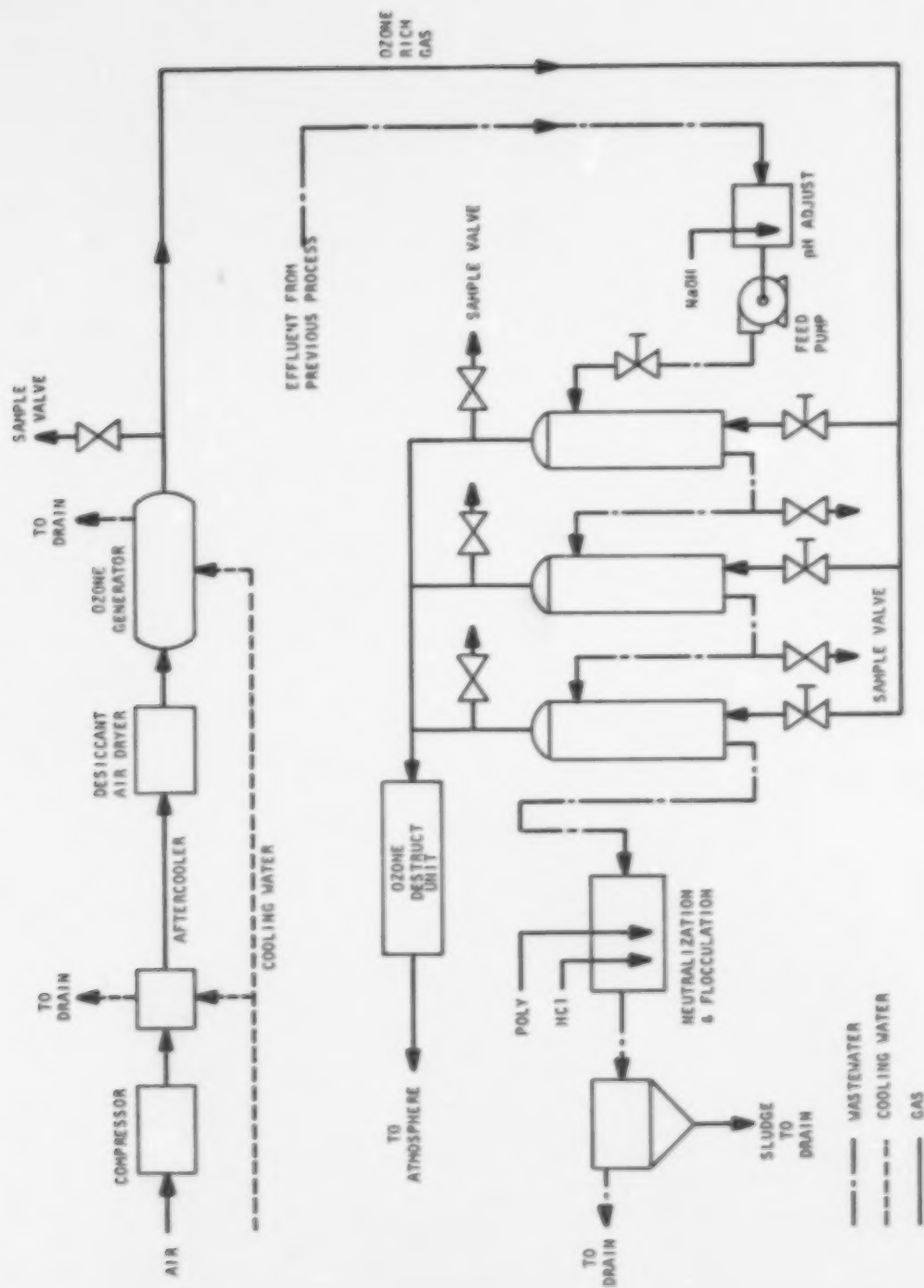


Figure 25. Schematic of pilot ozonation unit.



Figure 26-A. Ozonation pilot system - ozone generator.

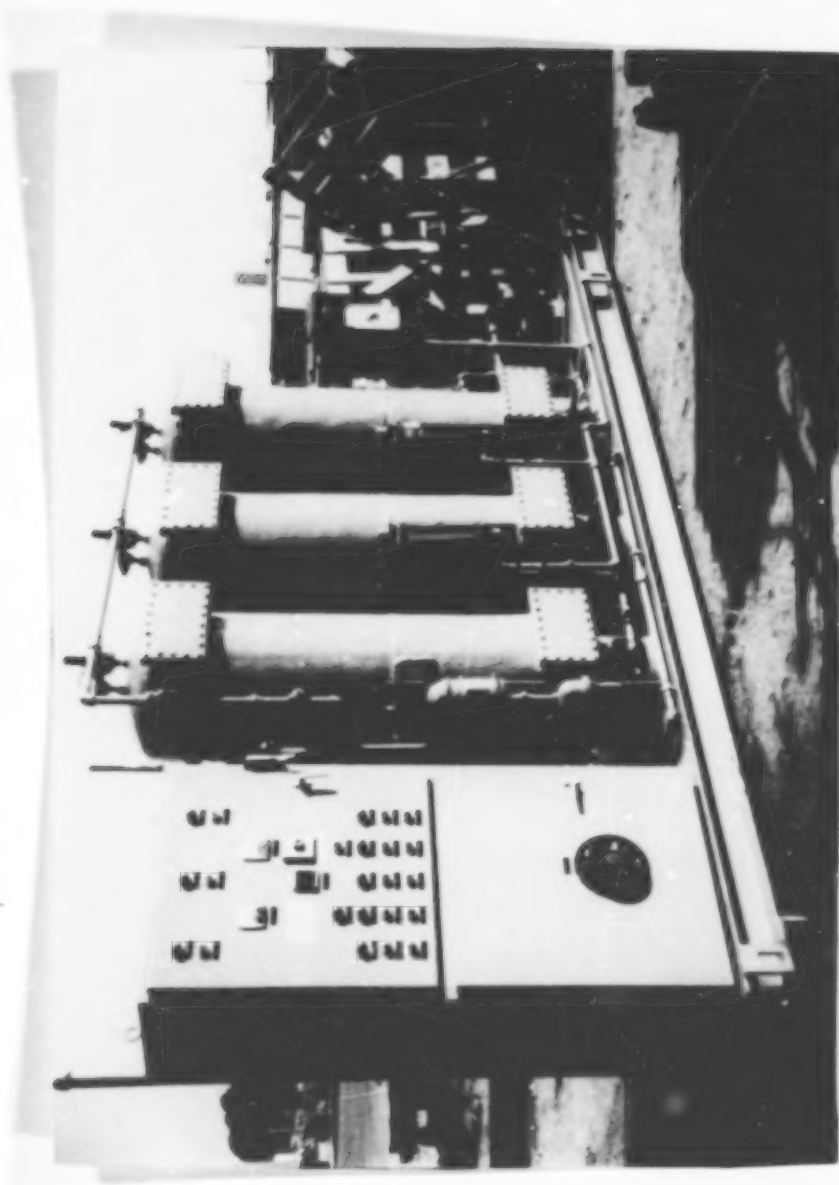


Figure 26-B. Ozonation pilot system - ozone contactors.

Experimental Program

The ozonation test program consisted of two separate sections, 1) ozonation of the raw wastewater after filtration and 2) ozonation of reverse osmosis (RO) brine. Operational variables investigated were hydraulic retention time (contact time) and ozone dosage.

The bench scale tests on the blast furnace raw wastewater indicated that ammonia would be the controlling factor in meeting BAT limitations. BAT limitations specify a maximum pollutant discharge in kg/1,000 kg of product (pounds/ton of product). Therefore, an effluent ammonia concentration that did not exceed BAT limitations for the specific research site was set as the objective of the test program for each wastewater investigated.

Filtered Wastewater Ozonation--

Dual media filtered raw wastewater was used as the influent to the ozonation system in the first portion of the test program. It was desired to determine various contact times and ozone dosages that produced an effluent ammonia concentration that bracketed the average ammonia concentration allowed by BAT limitations for the specific research site. This would allow the making of a rough estimation of contact time and ozone dose required for steel mills with different production capacities than the specific research site.

The sequence of events in a typical run was:

1. Start Emerzone system and chemical feed system. Set desired wastewater flow rate and ozone dose.
2. Allow system to reach equilibrium by running for a minimum of one contact time period.
3. Measure the ozone concentration in contactor feed and off-gases so that the concentration of ozone utilized could be determined.
4. Sample and composite the system effluent over a selected time period and obtain grab samples through system.
5. Shut down Emerzone system.

On-site measurement of ammonia was performed using an Orion selective-ion electrode for process operational checks. Daily composites and selected grab samples were preserved and shipped to the Rexnord Environmental Research Center laboratory for analysis. Samples for BOD₅ were sent to a local laboratory for analysis.

Measurement of the ozone concentration in the feed and off-gases was performed using absorption with a 1 percent potassium iodide (KI) solution and titration with 0.025N sodium thiosulfate (Na₂S₂O₃) and starch indicator as given in Standard Methods (2).

A typical data sheet used to record operational parameters is presented in Figure 27.

Date							
Run No.							
Feed Source							
Influent flow, mg/l							
Influent temperature, °F							
O ₃ dose, mg/l							
Time							
Cumulative time							
Compressor air flow, scfm							
Ozonator power setting, %							
O ₃ gas temperature, °F							
O ₃ gas pressure, psi							
Atmospheric pressure, mm Hg							
O ₃ gas concentration, mg/l							
Gas flow, scfm							
Contactor No. 1							
Contactor No. 2							
Contactor No. 3							
Off-gas O ₃ concentration, mg/l							
Contactor No. 1							
Contactor No. 2							
Contactor No. 3							
pH							
Before adjustment							
After adjustment							
Effluent							
Effluent NH ₃ concentration, mg/l							
NaOH % concentration							
Initial volume, gal							
Final volume, gal.							
HCl % concentration							
Initial weight, lbs							
Final weight, lbs							
Sample ESD No.							

Figure 27. Ozone data sheet.

RO Brine Ozonation--

The test program for the RO brine ozonation study was similar to the filtered raw wastewater program except for some limitations dictated by the 3.0 l/min (0.8 gpm) brine production of the RO system. Brine storage in a 1,700 l (450 gal.) tank permitted the use of reasonable contact times. However, the number of runs allowable in a day and the length of time per run were dictated by the availability of an adequate supply of brine.

Again, the bench scale tests indicated that the effluent ammonia concentration would be the controlling factor in meeting BAT limits. A range of effluent ammonia concentrations was developed so that the results of this study could be extrapolated to other steel mills.

ACTIVATED CARBON

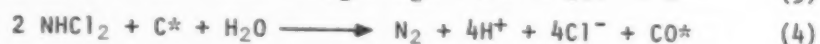
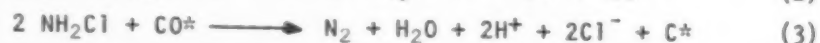
Introduction

Activated carbon adsorption systems were utilized in two of the four pilot system treatment trains investigated for blast furnace treatment. The activated carbon was used to remove residual chlorine from the alkaline chlorination process, which preceded the carbon in each treatment train.

Activated carbon is a well known dechlorinating agent for water. The reaction of free chlorine with activated carbon is illustrated in equation 1.



where C^* is an active site on the carbon surface and CO^* is a surface oxide. Reactions of mono- and dichloramines (combined chlorine residual) with activated carbon are illustrated by equations 2, 3, and 4



The ratio of chlorine to NH_3 and the pH of the wastewater determine the relative quantities of mono- and dichloramines. The reaction rate of free chlorine with activated carbon increases with decreasing pH due to the relative abundance of $HOCl$ compared to OCl^- , and increases with temperature. The rate of dichloramine removal by activated carbon depends on the relative abundance of mono and dichloramines. This is due to the fact that dichloramines react much faster than monochloramines. Dichloramines also react faster with activated carbon than $HOCl$.

Removal of free and combined chlorine also depends on many factors other than the relative abundance of free and combined species. Competition for sites on the carbon surface by chlorine residuals, organics, and suspended solids reduces the rate of chlorine removal. Pretreatment of the wastewater to remove suspended solids will significantly reduce their

impact on activated carbon. The organic and suspended solids reduce carbon surface area and plug carbon pores.

The size of the granular activated carbon will also affect the rate of residual chlorine removal. The surface reaction of all chlorine species is much faster than the diffusion of these species to the carbon surface. Smaller particles will then minimize diffusion time and increase chlorine removal.

Contact time for chlorine removal is normally between five to ten minutes. As the carbon surface is saturated with oxides, organics, and suspended solids, a five to ten minute contact time may not be sufficient to reduce the chlorine concentration to an acceptable level (1 mg/l). When the carbon is saturated and plugged to the point where the reaction rate is not high enough to reduce the chlorine to this acceptable level in the given retention time, the activated carbon must be removed from the column.

To effectively design an activated carbon system, reliable data must be obtained to adequately size the adsorbers. The data are best obtained by running the wastewater to be treated through a pilot scale system. Analysis of the test results are then used to obtain full-scale design criteria. In this study the concentration of chlorine in the feed to the activated carbon was dependent on the chlorine dose and chlorine demand in the alkaline chlorination process.

Description of Pilot Equipment

The mobile research trailers were equipped with two granular activated carbon adsorption systems: (1) a series of four - 35.6 cm (14 in.) diameter columns that were designed to remove chlorine residual from clarified, alkaline chlorinated raw water, and, (2) one 7.6 cm (3 in.) inside diameter column that treated chlorine residual from alkaline chlorinated reverse osmosis brine.

The four carbon columns were constructed of 35.6 cm (14 in.) diameter standard pipe. Influent wastewater was pumped into the bottom of the carbon columns by a self-priming pump. The flow was distributed over the carbon column by three diffusers. The carbon adsorption system was designed to operate in either the upflow or downflow mode, but for chlorine removal the upflow mode was chosen. Sample taps were located every 30.5 cm (12 in.) along the column, but samples were only taken at sample ports located 30.5 cm (1.0 ft), 91.5 cm (3 ft), and 152 cm (5 ft) from the bottom of the carbon bed. Pressure gauges are mounted at the top and bottom of each column to monitor headloss across the carbon bed. A schematic diagram of the four column adsorption system is illustrated in Figure 28.

The 7.6 cm (3 in.) diameter carbon column was constructed of 0.3 cm (1/8 in.) plexiglas and was five feet tall. The column was constructed to operate in the upflow or downflow mode. The upflow mode was chosen for this study.

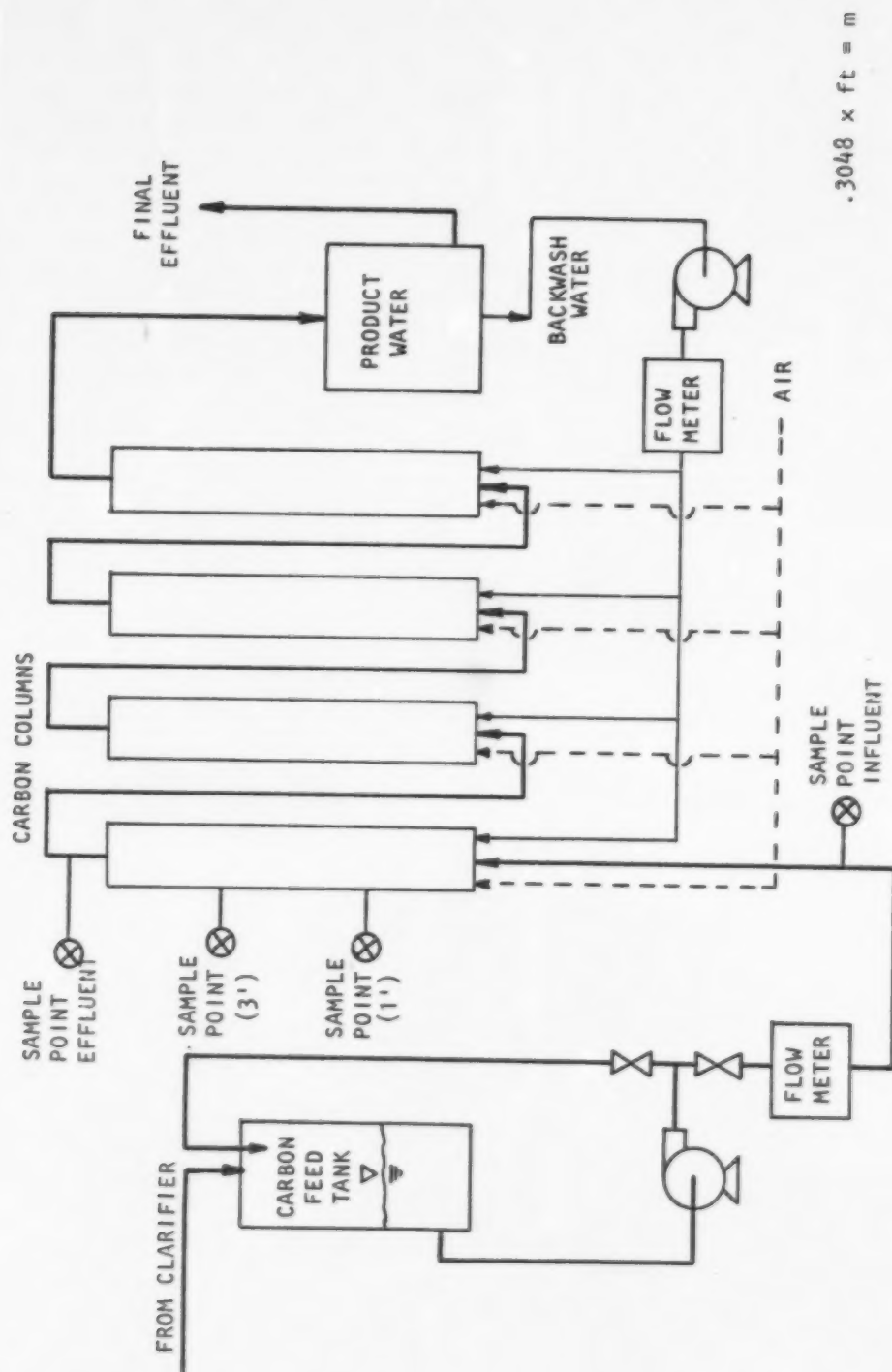


Figure 28. Granular carbon column test system.

Alkaline chlorinated reverse osmosis brine was pumped through the column by a self-priming ISCO pump. Flow was controlled by adjusting a clamp on the feed line and was measured using a bucket and stopwatch. Sample ports were located at the influent, 30.5 centimeter (one foot), 91.5 centimeter (3 foot), and effluent level. A schematic diagram of the 7.6 centimeter (3 in.) diameter carbon adsorption systems is in Figure 29.

Experimental Program

The activated carbon adsorption tests were designed to obtain operating data and design criteria for a full-scale dechlorination system. The dechlorination step was required to reduce free and combined chlorine residuals from the alkaline chlorination process. Two treatment trains investigated had alkaline chlorination units.

Flow rates and carbon dosages were calculated to provide test sequences of sufficient duration to obtain data which would illustrate a decrease in the carbon's capacity to reduce chlorine. The data provided design criteria to determine expected bed life.

The 35.6 cm (14 in.) diameter columns were used in the treatment train illustrated below.

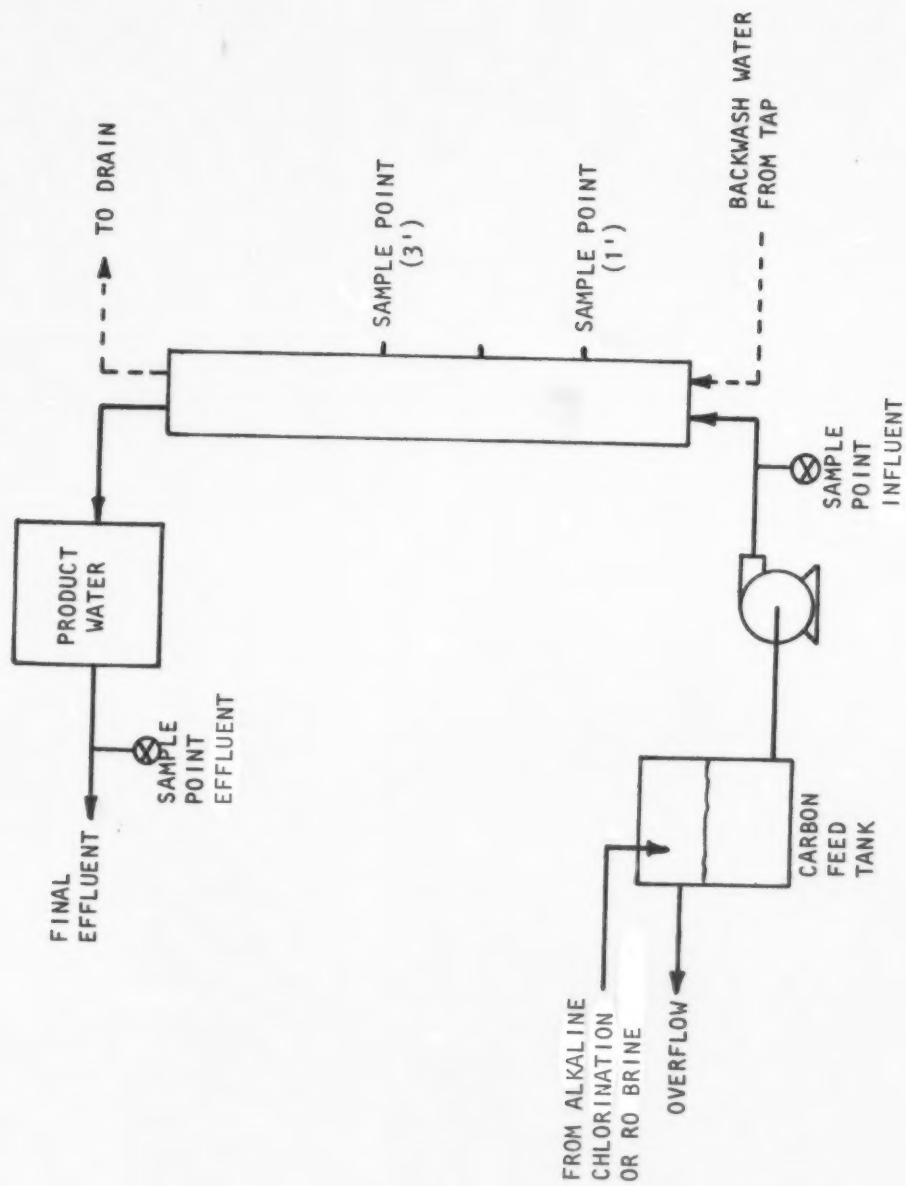
raw wastewater → alkaline chlorination → clarification → activated carbon → effluent

Even though the test trailers were provided with four columns, only one was required for this particular study. The operating parameters for the 35.6 cm (14 in.) diameter columns are outlined below.

- Carbon mesh, 8 x 30
- Carbon type, bituminous
- Carbon weight per column, 67 kg (147 lbs)
- Bed depth, 152 cm (60 in.)
- Linear flow rate, $0.163 \text{ m}^3/\text{day}/\text{m}^2$ (4 gpm/ft)
- Contact time, 10 minutes/column
- Dimensions of carbon column
 1. Column diameter, 35.6 cm (14 in.)
 2. Column height, 229 cm (90 in.)

The 7.6 cm (3 in.) diameter column was used to reduce chlorine in the treatment train illustrated below.

reverse osmosis brine → alkaline chlorination → adsorption → effluent



.3048 x ft = m

Figure 29. Small granular carbon column test system.

The 7.6 cm (3 in.) diameter column was operated in an upflow mode. The operating parameters are outlined below.

- Carbon mesh, 8 x 40
- Carbon type, bituminous
- Carbon weight in column, 1.684 kg (3.7 lbs)
- Bed depth, 96.5 cm (38 in.)
- Linear flow rate, $0.10 \text{ m}^3/\text{day}/\text{m}^2$ (2.46 gpm/ft²)
- Volumetric flow rate, 466 ml/min. (0.123 gpm)
- Contact time, 9.6 minutes
- Dimensions of carbon column,
 1. Column diameter, 7.6 cm (3 in.)
 2. Column height, 152 cm (60 in.)
 3. Cross sectional area, 46.45 cm^2 (7.2 in.²)

Sampling Procedures

Grab samples were taken periodically from sample ports on both columns. These samples were analyzed on site for pH, temperature, turbidity, and Cl_2 concentration. The chlorine analysis method used was the Iodometric Method which is outlined in Standard Methods for the Examination of Water and Wastewater, 14th Edition (2).

Composite samples were taken every hour for 24 hours from the 35.6 cm (14 in.) column, or over the length of the test run, whichever was shorter. These samples were preserved accordingly and shipped back to the Rexnord Environmental Research Center analytical lab for analysis. Composite samples were tested for suspended solids, volatile suspended solids, color CN_T , CN_A , Phenol, NH_3 , TKN, $\text{NO}_3\text{-NO}_2$, BOD, COD, sulfide, fluoride, TDS, chloride, and total phosphorous. Grab samples were also collected for oil and grease analyses. Routine samples were also taken several times daily on site for pH, temperature, turbidity and chlorine residual. Included in this group of parameters are the seven pollutants which are limited by BAT effluent guidelines for blast furnace wastewater discharge.

A copy of the data sheet used to record routine operating data is shown in Figure 30.

Research Site: BURNS HARBOR

Media Type: BITUMINOUS CARBON

Treatment Train: _____

Date: _____ Operator: _____

System: _____

Time: _____

Flow Rate, gpm _____

Totalizer Reading, gal. _____

Influent

Effluent

Sampling

Time: _____

Water Meter Reading: _____

Column Pressure

P In: _____

P Out: _____

Source	Temp.	pH
Influent		
Effluent		

Source	Sample Point	Concentration
Influent	No. _____	_____
Mid-depth	No. _____	_____
	No. _____	_____
Effluent	No. _____	_____

Backwash

Time: _____

BEFORE	
P _{in}	P _{out}

Flow	Time

AFTER	

Figure 30. Activated carbon operational data sheet.

SECTION 7

BLAST FURNACE DATA EVALUATION

INTRODUCTION

This section presents a summary of the findings at the blast furnace research site. A results section is provided for each of the technologies investigated (alkaline chlorination, chemical treatment, magnetic filtration, dual media filtration, reverse osmosis, ozonation and activated carbon). Each technology section includes a presentation and discussion of results, a full-scale system description, and full-scale system cost and space requirements.

The field study consisted of an extensive sampling and analytical program. Analytical results of the major parameters tested are presented in Appendix Table A-2. The other parameters tested (SO_4^{2-} , Cl_2 , etc.) are discussed in the respective sections. A summary of the composite sample analytical results is presented in Table 3. The blast furnace blowdown feed to the pilot system remained relatively stable during the testing period except for a brief interval during mid-February, 1979. During this period, one of the blast furnace thickeners was removed from service, causing abnormally high concentrations for some of the parameters being tested in the blowdown effluent. Ammonia averaged 38.9 mg/l during the study with a range from 17.0 to 72.0 mg/l. The average cyanide-A concentration was <0.16 during the study with a range from <0.01 to 0.68 mg/l. Total cyanide approximated cyanide-A concentrations throughout the study. Suspended solids ranged from 26 mg/l to 3,595 mg/l and averaged 237 mg/l. The treatment efficiencies of each technology investigated are discussed in each respective results section.

Bench tests performed on the blast furnace blowdown concluded that the most promising treatment methods for meeting BAT limitations were chlorination, ozonation, and reverse osmosis. Also, chemical treatment, clarification, filtration and activated carbon (if required for dechlorination prior to discharge) were found to be essential pretreatment and/or post-treatment adjuncts.

Each treatment was operated to meet the following effluent concentration goals as shown in the EPA Effluent Guidelines and Standards for Iron and Steel Manufacturing; Subpart D - Blast Furnace (Iron Subcategory) BAT limitations (Table 4):

TABLE 3. BLAST FURNACE RESEARCH SITE RANGE
OF PARAMETER VALUES DURING TESTING

		pH	SS mg/l	NH ₃ as N mg/l	CN _A mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l
Raw Wastewater	Max.	8.30	3595	72.1	0.68	1.10	21.4	0.3
	Min.	7.10	26	17.0	<0.01	0.008	6.44	<0.1
	Avg.	--	237	38.9	<0.16	0.060	12.3	<0.2
Ozonated Filter Effluent	Max.	7.80	16	34.0	0.50	0.960	16.2	0.3
	Min.	6.90	1	3.1	<0.01	0.001	10.8	<0.1
	Avg.	--	8	16.4	<0.08	0.226	12.9	<0.2
Alkaline Chlorination Clarified Effluent	Max.	7.40	37	11.9	11.0	0.386	20.8	<0.2
	Min.	6.30	7	0.48	<0.01	<0.001	10.5	<0.1
	Avg.	--	23	5.63	<0.02	<0.033	15.0	<0.2
Activated Carbon Effluent	Max.	7.45	44	15.9	0.06	0.545	21.1	<0.2
	Min.	6.60	6	0.65	<0.01	0.002	10.0	<0.2
	Avg.	--	18	7.61	<0.02	0.118	15.2	<0.2
DMF Effluent (chemical pretreatment)	Max.	7.90	57	52.0	0.27	0.176	10.9	<0.2
	Min.	6.50	3	4.5	<0.01	0.008	3.70	<0.2
	Avg.	--	15	36.4	<0.04	0.040	7.76	<0.2
RO Product	Max.	8.20	6	30.4	<0.01	0.151	0.80	<0.2
	Min.	5.35	1	3.3	<0.01	<0.001	0.28	<0.2
	Avg.	--	3	17.5	<0.01	<0.061	0.54	<0.2
RO Brine	Max.	7.50	29	172	0.51	0.660	35.2	<0.2
	Min.	6.50	26	60.0	<0.01	0.011	6.60	<0.2
	Avg.	--	28	114	<0.08	0.276	17.3	<0.2
Ozonated RO Brine	Max.	9.90	78	91.5	0.01	0.056	27.6	--
	Min.	6.30	31	39.0	<0.01	0.002	9.10	--
	Avg.	--	52	57.8	<0.01	0.018	18.1	--
Alkaline Chlorinated RO Brine	Max.	7.00	88	23.8	0.03	0.018	21.9	<0.2
	Min.	5.80	34	<0.05	<0.01	0.002	8.76	<0.2
	Avg.	--	51	<9.91	<0.01	0.009	15.5	<0.2
Activated Carbon Effluent (Alkaline Chlorinated RO Brine)	Max.	8.10	42	30.0	0.04	0.137	20.9	<0.2
	Min.	6.60	19	2.50	<0.01	<0.001	9.5	<0.2
	Avg.	--	32	16.4	<0.02	<0.035	13.9	<0.2

TABLE 4. IRON SUBCATEGORY - BAT EFFLUENT GUIDELINES

Parameter	Effluent guidelines (kg/1000 kg product)
TSS	0.013
CN-A	0.00013
Phenol	0.00026
NH ₃	0.0052
Sulfide	0.00016
F	0.0104
pH	6.0-0.0

Each day during the entire pilot program, production information was obtained from the steel mill personnel. This information was recorded on a data sheet as shown in Figure 31. Blowdown readings were taken daily by Rexnord field engineers at 11:00 p.m. This time was selected so that daily blowdown readings coincided with the blast furnace production day, which also runs from 11:00 p.m. to 11:00 p.m..

The above information was vital to the field engineers who used the data to determine daily BAT discharge concentrations for the blast furnace blowdown. Field test runs were then planned to reflect changes in the BAT allowable concentrations. As an example, assume the following:

Day 1: Iron ore production = 10,884 metric tons (12,000 tons)
Blowdown = 3,785 m³/day (1 mgd)

From BAT effluent guidelines, average ammonia allowed in the effluent is 0.0052 kg/1000 kg production (0.0052 lb/1000 lb production) to meet BAT requirements.

Then:

$$\begin{aligned}
 &0.0052 \text{ kg/1000 kg production} \times 1000 \text{ kg/metric ton} \\
 &\quad \times 10,884 \text{ tons} \\
 &= 56.66 \text{ kg (124.8 lb) NH}_3 \text{ allowed/day} \\
 &\text{mg/l NH}_3 \text{ allowed} = \frac{56.66 \text{ kg}}{.001 \cdot 3,785} = 15 \text{ mg/l}
 \end{aligned}$$

Similar allowable concentrations were developed for each of the BAT parameters and this information is shown in Appendix Table A-3. The information developed was plotted daily in the research trailers to serve as a planning guide for the field engineer.

During the reverse osmosis study, the membrane rejected significant concentrations of BAT parameters. If a 75 percent recovery and reuse of the product water is assumed, a relationship can be developed for treatment and ultimate discharge of the concentrated brine. For example, assume the following:

Day 1: Iron ore production = 10,884 metric tons (12,000 tons)
Blowdown = 3,785 m³/day (1 mgd)
RO Brine Flow (75% recovery) = 946 m³/day (0.25 mgd)

From previous calculations, the allowable BAT discharge for ammonia is 56.66 kg (124.8 lb) allowed/day.

Then:

$$\text{mg/l NH}_3 \text{ allowed} = \frac{56.66 \text{ kg}}{.001 \cdot 946} = 60 \text{ mg/l}$$

Thus, because of the lower flow associated with treatment of the reverse osmosis brine, allowable concentrations are four times that allowed for the raw wastewater.

LABORATORY QUALITY CONTROL

The Rexnord laboratory performed a series of quality control tests to validate the accuracy of the analyses performed on the samples collected at the research site during the study period. A summary of these quality control tests is presented in Table 5. The average standard deviations of duplicate measurements for the various parameters were grouped in concentration ranges to illustrate the change in precision with concentration. Percent recoveries of spiked samples were determined for several parameters. It is important to note that the percent recoveries given for sulfide and cyanide amenable to chlorination were quite good. This indicated that the methods of analysis for these parameters were in control and that very low concentrations found during the study are reliable.

DESCRIPTION OF COST ASSUMPTIONS

Delivered costs for the various principal pieces of equipment were obtained from vendor quotations and from a recent edition of the Dodge Manual for Building Construction Pricing and Scheduling (6). The total plant cost includes all of the funds required to procure equipment, land and buildings in addition to funds required to design, build and bring the facility to acceptable operation. To determine the total plant cost, the Chilton method (7) of cost estimation was utilized. The Chilton method uses a

TABLE 5. SUMMARY OF STATISTICAL EVALUATION
OF LABORATORY QUALITY CONTROL ANALYSIS

Parameter	Concentration range mg/l	Standard deviation	Percent recovery
SS	0 - 30	2 (138)	
	31 - 100	4 (95)	
	101 - 350	14 (55)	
	351 - 2000	51 (56)	
	>2000	900 (63)	
VSS	0 - 10	2 (71)	
	11 - 30	3 (45)	
	31 - 500	22 (48)	
	501 - 1000	74 (12)	
	>1000	154 (15)	
TDS	0 - 100	1 (5)	
	101 - 1000	25 (15)	
	1001 - 3000	30 (97)	
	>3000	78 (38)	
CN _A	0.00 - 0.05	0.00 (2)	
	1.10 - 1.33		101
CN _{TOTAL}	0.00 - 00.7	0.00 (2)	
NH ₃	0.0 - 5.9	0.1 (17)	
	6.0 - 20.9	0.2 (15)	
	21.0 - 50.0	0.7 (25)	
	>50.0	1.2 (2)	
	4.06 - 4.98		96.5
	20.0 - 54.0		101.1
TKN	0 - 10.9	0.5 (14)	
	11.0 - 25.9	0.5 (9)	
	26.0 - 50.0	1.1 (17)	
	>50.0	0.7 (2)	
	0.53		107
	3.20		98
S	0.0 - 0.2	0.0 (4)	
	0.0 - 1.0		99.8
F	0.0 - 35.0	0.5 (31)	
	0.70		91.4
	21.1 - 27.3		90.8
TOC	0 - 14	0 (3)	
Cl	0 - 500	4 (7)	
	501 - 1000	9 (2)	
	1001 - 1500	27 (13)	
	>1500	53 (2)	
	610 - 1560		96.6
SO ₄	89 - 198	2 (3)	

() Indicates number of standard deviations of duplicates used
to obtain average standard deviation

factored approach. That is, the delivered costs of the principal pieces of equipment in the system are summed and then multiplied by various factors appropriate for the type of system being built. Factors for most items have a selection range which requires the judgement of the estimator to make an appropriate selection within the range.

Similar factors were selected for all systems which permits consistent comparison between systems.

The particular Chilton factors used for the various treatment methods and a costing example are listed in Tables A-4 and A-5 in the Appendix.

The principal items that are included in the cost estimates are the following.

1. Capital Equipment.
2. Buildings.
3. Land Acquisition and Site Preparation.
4. Utilities and Yard Services (air, water, electricity, etc.).
5. Engineering Services.
6. Contractors Fees.
7. Contingencies.

The assumptions that were used in preparing the cost estimates are listed in Table 6.

As shown in Table 6, costs were developed for six systems, three each for the raw wastewater and the reverse osmosis brine. The design flows selected were representative of typical blast furnace blowdown discharges.

The next sections in this report will discuss the testing results of each process element investigated from a standpoint of performance, costs, and space requirements.

RESULTS OF THE ALKALINE CHLORINATION STUDY

Introduction

Alkaline chlorination was investigated to reduce influent concentrations of ammonia, cyanide, and phenol to within Best Available Technology (BAT) Economically Achievable limits. The pilot test procedure consisted of passing the wastewater through a series of completely mixed reaction tanks. In the first chamber, sodium hydroxide and sodium hypochlorite were added to oxidize the cyanide present. In the second chamber, hydrochloric acid and ferric chloride (a coagulant) were added to neutralize the waste. This chamber was used for ammonia removal. Waste exiting this chamber was dosed with polymer and settled in a clarifier prior to being dechlorinated on activated carbon. As a byproduct of treatment, other parameters exerting a chlorine demand (phenol, etc.) were also oxidized.

TABLE 6. ASSUMPTIONS USED TO DEVELOP COST AND DESIGN
FIGURES FOR FULL SCALE SYSTEMS

1. Electric Costs: \$0.03/KWH		
2. Capital Amortization: 6.875 percent 20 year service life		
3. Design Flow:		
	Raw Wastewater	Reverse Osmosis Brine
	950 m ³ /day (0.25 mgd)	237 m ³ /day (0.0625 mgd)
	2,840 m ³ /day (0.75 mgd)	710 m ³ /day (0.1875 mgd)
	5,680 m ³ /day (1.5 mgd)	1,419 m ³ /day (0.375 mgd)
4. Labor: \$18,000/year skilled \$16,000/year unskilled		
5. The costs developed are exclusive of supervisory and laboratory expenses, personnel overhead, fringe benefits, indirects (interest on capital during construction, etc.), sludge handling and disposal, etc.		
6. All prices are effective as of May 1, 1979.		

The mobile treatment facilities were equipped with two independent alkaline chlorination units, one for treating the raw blowdown and a smaller system for evaluating treatment of the reverse osmosis brine. Descriptions of this test equipment have been given previously in Section 6.

Alkaline chlorination tests on the blast furnace scrubber blowdown were performed from January 30 to February 23, 1979. During this study period, 20 test runs were conducted. Treatment of the reverse osmosis brine by alkaline chlorination was investigated from March 19 to 26, 1979. A total of 9 runs were performed during this phase of testing.

Alkaline Chlorination Of The Blast Furnace Scrubber Blowdown

Tables B-1 and B-2 in the Appendix present the data analysis of all samples taken during the alkaline chlorination of the blast furnace scrubber blowdown. Included in these tables are information on flow rates investigated, influent and effluent pollutant concentrations, chemical usage rates, etc. The most pertinent information has been sifted and carried forth to the summary presented in Table 7. This table shows that alkaline chlorination removed significant quantities of cyanide-A and ammonia nitrogen. Influent versus effluent ammonia for each of the twenty runs is illustrated in

TABLE 7. BLAST FURNACE WASTEWATER
ALKALINE CHLORINATION DATA

Run No.	Flow gpm	Cl ₂ :NH ₃ ratio	Ammonia, mg/l		Cyanide A, mg/l	
			Inf.	Eff.	Inf.	Eff.
1	3.0	14.4:1	46*	1.3*	-	-
2	3.0	13.9:1	38.3	1.7	0.12	<0.01
3	3.0	12.0:1	37.5*	1.0*	-	-
4	3.0	7.6:1	35.9	7.4	0.06	<0.01
5	3.0	9.6:1	36.3	4.8	0.17	<0.01
6	5.0	10.0:1	32.9	0.48	0.19	<0.01
7	5.0	7.3:1	29.0	0.67	0.10	<0.01
8	5.0	15.2:1	29.2	0.67	0.02	<0.01
9	5.0	6.9:1	30.4	10.3	0.01	<0.01
10	5.0	8.0:1	32.4	5.2	0.02	<0.01
11	5.0	7.4:1	46.0*	6.95	0.18	<0.01
12	5.0	8.6:1	32.2	10.9	0.43	<0.01
13	5.0	9.6:1	28.6	5.7	0.10	<0.01
14	6.4	7.1:1	32.8	5.2	0.12	0.02
15	8.5	8.8:1	33.3	7.35	0.34	0.06
16	8.5	7.6:1	34.0	6.15	0.23	<0.01
17	8.5	7.9:1	38.4	11.9	0.28	<0.01
18	8.5	11.5:1	37.8	4.8	0.23	<0.01
19	8.5	8.0:1	34.5*	3.1*	-	-
20	8.8	7.1:1	32.0*	16.0*	-	-

Note: * Based on field measurements
gpm x 3.785 = l/min

Figure 32. The pilot study confirmed the bench test findings that if the influent ammonia concentrations were reduced below BAT limitations, other oxidizable parameters (cyanide, phenol, sulfide) were also reduced to acceptable levels. Chlorination also significantly decreased TOC and organic nitrogen levels. The data shown in Table 7 indicate that as the chlorine to ammonia ratio was increased, ammonia removal improved. This is shown graphically in Figure 33. Ammonia concentrations remained high until $\text{Cl}_2:\text{NH}_3$ ratios exceeded 7.3:1 and then dropped sharply as this ratio was increased. Chlorine to ammonia ratios from 6.9:1 to 15.2:1 were studied. Effluent ammonia concentrations as low as 0.48 mg/l were achieved, indicating that breakpoint had occurred. Excellent cyanide-A removals were observed at all $\text{Cl}_2:\text{NH}_3$ ratios investigated. During the study, flow rates of 11.36 l/min (3 gpm), 18.93 l/min (5 gpm), and 32.2 l/min (8.5 gpm) were investigated. One run each was also conducted at 24.2 l/min (6.4 gpm) and 33.3 l/min (8.8 gpm). The results indicated that good ammonia removals could be obtained at all flow rates studied, provided the $\text{Cl}_2:\text{NH}_3$ ratio was high enough.

Raw feed water quality (blast furnace scrubber blowdown) remained relatively stable during the testing period with the exception of suspended solids. As shown below in Table 8, ammonia feed concentrations during the testing period varied from 17.0 mg/l to 46.0 mg/l with an average of 33.4 mg/l.

TABLE 8. ALKALINE CHLORINATION OF BLAST FURNACE SCRUBBER BLOWDOWN - INFLUENT QUALITY DURING PILOT TESTING

	Temp. °C	pH	SS, mg/l	NH, mg/l	CN-A, mg/l	Fluoride, mg/l	Phenol, mg/l	Sulfide, mg/l
Average	22.2	-	663	33.4	<0.16	15.33	0.027	<0.2
Maximum	26	8.0	3595	46.0	0.43	21.4	0.077	0.29
Minimum	16	7.3	26	17.0	<0.01	11.5	0.011	<0.1

The average cyanide-A in the influent was <0.16 mg/l with a range from <0.01 mg/l to 0.43 mg/l. Total cyanide values (shown in the Appendix Tables) closely approximated cyanide-A concentrations. Suspended solids showed the widest variation during the pilot study, ranging from a low of 26 mg/l to a high of 3,595 mg/l. The several days of high suspended solids that occurred, resulted from a shutdown of one of the blast furnace scrubber water thickeners. However, this temporary condition did not significantly affect influent ammonia and cyanide concentrations, although it did increase the solids loading to the pilot clarifier located downstream of the alkaline chlorination system. Wastewater temperature ranged from 16°C to 26°C with an average of 22.2°C. For a typical test run, a grab sample of the raw wastewater was analyzed in the field for ammonia concentration

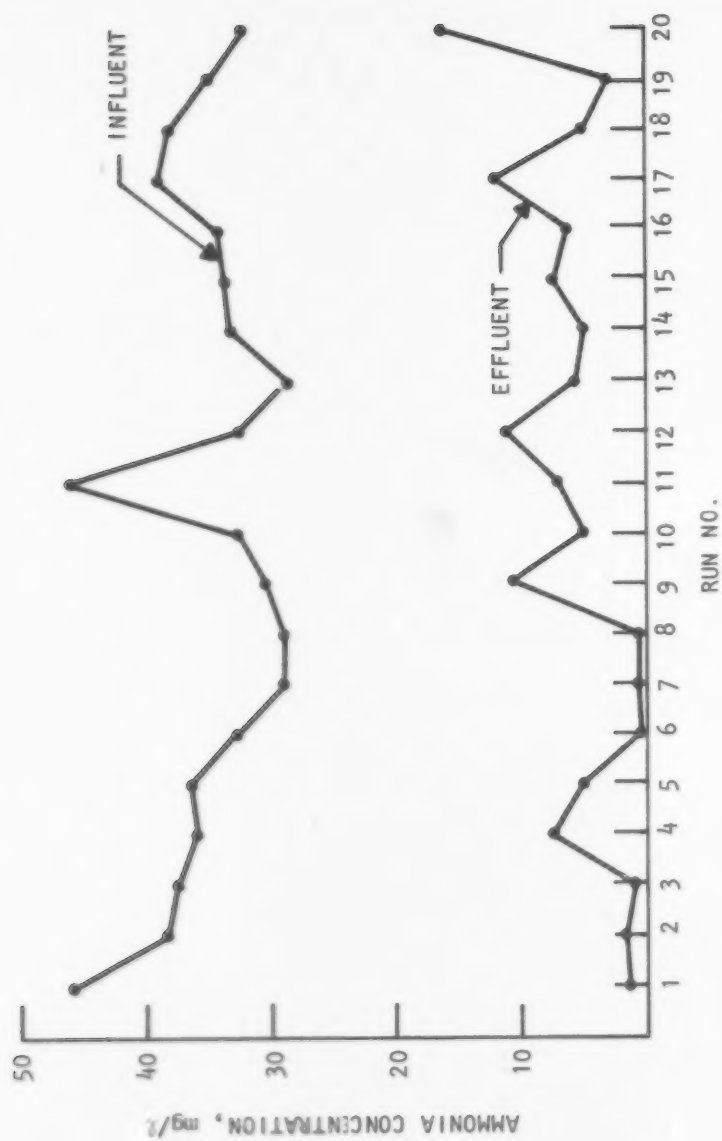


Figure 32. Influent vs. effluent ammonia concentration alkaline chlorination-raw wastewater.

NOTE: gpm x 3.785 = liters/min.

Δ - FROM FIELD MEASUREMENTS

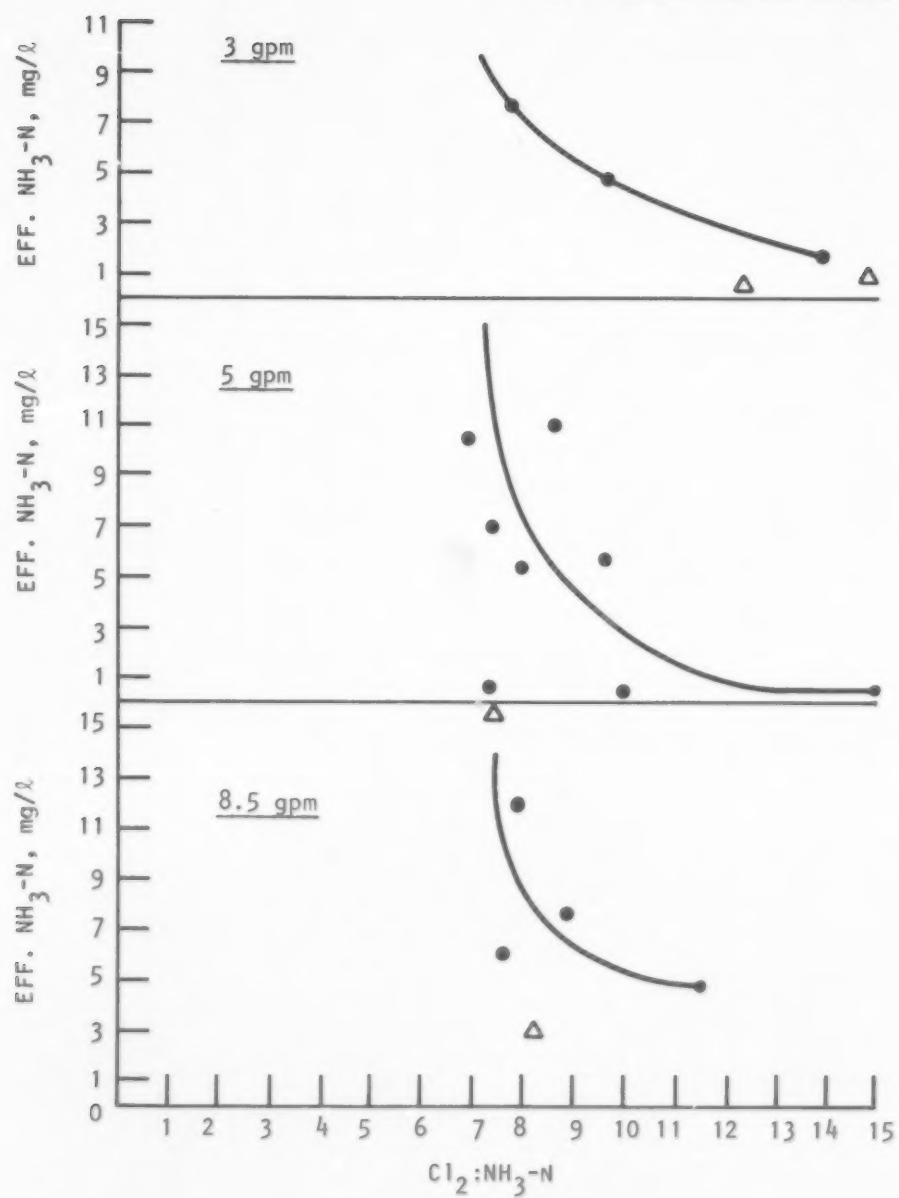


Figure 33. EFF. $\text{NH}_3\text{-N}$ VS. $\text{Cl}_2:\text{NH}_3$ ratio.

using an Orion ammonia probe. A chlorine to ammonia ratio was selected and the sodium hypochlorite feed rate was adjusted accordingly. The raw waste entered the first reaction chamber and was completely mixed to a pH of about 11.0 (using a pH controller and a sodium hydroxide solution). The sodium hypochlorite was also added to this chamber. As the cyanide is converted to cyanate, cyanogen chloride is formed as an intermediate step. This is extremely irritating, even in low concentrations. A Dreager gas detector tube was available for routine measurement of airborne cyanogen chloride. The research trailers were equipped with exhaust fans, and no detectable concentrations of this gas were produced during the study. Detention times in the cyanide conversion tank ranged from 6.8 minutes to 20 minutes.

The waste then passed into the second completely mixed reaction chamber where the pH was reduced to about 7.0 using a hydrochloric acid feed and a pH controller. Ferric chloride was also added to this compartment to take advantage of its acidity. The ferric chloride was used as a coagulant downstream in the clarification process. Typically, 100 mg/l were added to the wastewater. Following neutralization, the waste entered a small completely mixed tank where pH was continuously monitored and controlled using a feedback loop to the previous tank. Exiting water was contacted with polymer added in-line and settled in a clarifier. A polymer dosage of 0.5 mg/l was used throughout the study. The combined contact time in the two neutralization tanks was from 8.2 to 24 minutes, depending on flow rate. Additional contact time was provided in the clarifier (from 83 to 240 minutes depending on flow rate).

At the start of the run, the influent sampler was iced and turned on. Measurements were made of chemical stock tank volumes, and chemical addition rates were checked for the sodium hypochlorite, ferric chloride, and polymer. Influent data were also recorded (pH, temperature, ammonia, Cl_2 residual). The system was allowed to come to equilibrium and the clarifier effluent sampler was iced and turned on. During equilibrium, grab samples were taken from the first and second reaction tanks for suspended solids, ammonia, and cyanide analysis. When the run was completed, chemical tank volumes were checked and recorded, and the samples were collected, preserved, and refrigerated. A new flow rate and $\text{Cl}_2:\text{NH}_3$ ratio was selected, and the next run was initiated. The samples were iced prior to being shipped back to the Rexnord Environmental Research Center in Milwaukee for analysis. Routine analysis included suspended solids, TOC, BOD, oil and grease, cyanide, phenol, ammonia, sulfite, fluoride, dissolved solids, nitrite/nitrate and TKN. Samples were also collected and analyzed on-site for pH, chlorine residual, temperature, and ammonia.

Values of pertinent parameters as they passed through the alkaline chlorination system are shown graphically in Figure 34 for a typical run (Run No. 18). For example, the influent pH for this run was 8.0 and was elevated to 11.9 in the cyanide reaction tank; then reduced to 7.7 in the ammonia reaction tank, and exited the clarifier at 7.1. The effluent pH was generally maintained at 7.0 or less to make the water more amenable to dechlorination on the downstream activated carbon.

The figure also shows that the alkaline chlorination process produced suspended solids, but they were effectively removed in the pilot clarifier with the aid of chemical coagulants. Figure 34 shows that a hydroxide floc formed when the pH was elevated above 11.0 for cyanide removal. This would appear to be an ideal time to precipitate any undesirable metals present in the waste. Following settling, the waste could then be neutralized to oxidize the ammonia. Although intermediate settling appeared feasible for this particular waste, this type of investigation was beyond the workscope of this project. The figure also illustrates the removals of cyanide and ammonia. Cyanide was efficiently destroyed in the first reaction chamber under alkaline conditions. Some ammonia reduction occurred initially with the reaction being completed as the waste passed through neutralization in the ammonia reaction tank and clarifier. Chlorine residual was highest in the first chamber, and decreased with time.

Although not shown graphically, alkaline chlorination increased dissolved solids in the waste due to the addition of chemical salts. During the testing period, influent dissolved solids averaged 1,207 mg/l while the average effluent concentration was 2,285 mg/l. This represents a 47 percent increase through the system. Analytical data were also taken periodically to monitor Total Kjeldahl Nitrogen (TKN) removal. On the average, approximately 82 percent of the TKN was removed.

A comparison of selected alkaline chlorination run effluents to BAT limitations is presented in Table 9. Analytical data for one run each at flow rates of 11.36 l/min (3 gpm), 18.93 l/min (5 gpm), and 32.2 l/min (8.5 gpm) have been shown for a moderate and high chlorine to ammonia ratio.

Comparing the maximum and average cyanide values found during the study (Table 8) to the BAT limitations presented in Table 9, it is apparent that average influent cyanide-A values (<0.16 mg/l) do not exceed allowable BAT values (0.271 mg/l). However, maximum pilot study cyanide concentrations (0.43 mg/l) would violate future restrictions, and a cyanide destruct system would be necessary. The methods used to develop the expected BAT values have been discussed in-depth previously in the Section 7 Introduction. The data in Table 9 indicate that required BAT effluent quality was satisfactorily met for all of the runs listed. In fact, BAT limits were typically met for all pilot test runs that had a $\text{Cl}_2:\text{NH}_3$ ratio in excess of 7.3:1. Based on the data presented in Table 9 and Figure 33, it appears that breakpoint chlorination is reached at a $\text{Cl}_2:\text{NH}_3$ ratio of about 10.1:1. For Run No. 6, this ratio resulted in an effluent ammonia concentration of 0.48 mg/l. Also, the lowest residual chlorine concentrations were achieved at $\text{Cl}_2:\text{NH}_3$ ratios of 9.5:1 to 10.0:1. A full-scale system operating at a chlorine to ammonia ratio of 10:1 could expect a total chlorine residual of about 40 mg/l. Additional investigations would be necessary to determine if $\text{Cl}_2:\text{NH}_3$ ratio could be slightly adjusted to produce a lower chlorine residual in the effluent. The chlorine was easily removed by activated carbon.

As mentioned previously, it would not be necessary to operate the alkaline chlorination system to breakpoint to reduce ammonia levels below BAT limits. Several studies have investigated the addition of only enough chlorine to

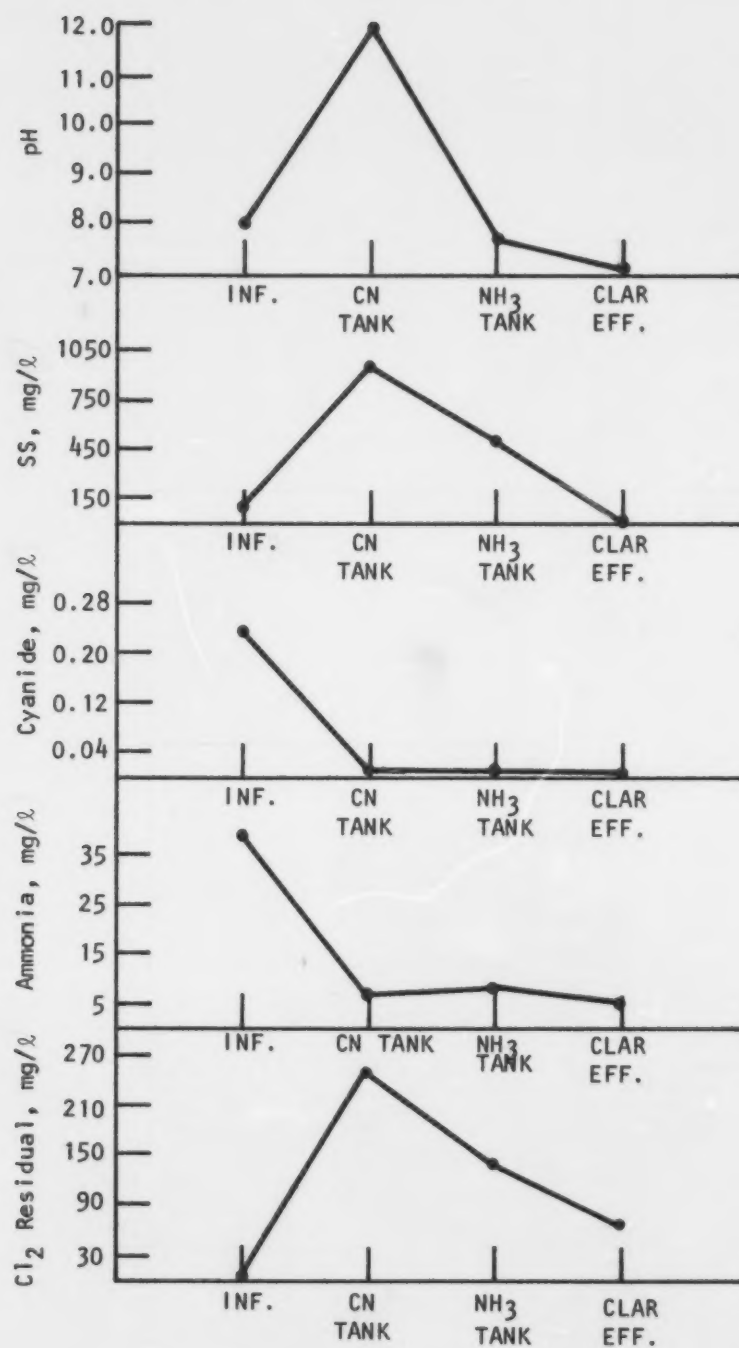


Figure 34. Values of various parameters through alkaline chlorination system.

TABLE 9. COMPARISON OF SELECTED ALKALINE CHLORINATION
RUN EFFLUENTS TO BAT LIMITATIONS

Run no.	Flow rate gpm	Cl ₂ :NH ₃ ratio	Ammonia, mg/l	Cyanide-A, mg/l	Fluoride, mg/l	Phenol, mg/l	Sulfide, mg/l	Suspended Solids, mg/l	pH
2	3.0	13.9:1	1.7	<0.01	10.5	0.014	<0.2	8	7.3
6	5.0	10.0:1	0.48	<0.01	12.5	0.014	<0.2	16	7.3
18	8.5	11.5:1	4.8	<0.01	16.9	0.005	<0.2	36	7.1
4	3.0	7.6:1	7.4	0.11	14.2	0.012	<0.2	11	6.7
10	5.0	8.0:1	5.2	<0.01	13.4	0.009	<0.2	25	6.7
16	8.5	7.6:1	6.15	<0.01	18.0	0.005	<0.2	25	7.4
BAT limitations			10.85	0.271	21.71	0.543	0.334	27	6.0-9.0

Note: gpm x 3.785 = l/min

oxidize the ammonia by forming monochloramines (8,9,10). Using this procedure, the monochloramines would be removed on activated carbon. If dechlorination were not used, large concentrations of total chlorine residual would be discharged.

There are two advantages to using this procedure for ammonia removal:

1. The theoretical chlorine to nitrogen ratio for removing 100 percent of the ammonia would be reduced from 7.6:1 down to about 5:1 for the formation of monochloramine according to the equation.



During the pilot study, an actual $\text{Cl}_2:\text{NH}_3$ ratio of about 10:1 was found necessary for breakpoint.

2. There would be a reduction in the alkalinity requirement (assuming Cl_2 gas feed) and dissolved solids formation.

The monochloramine removal would be by activated carbon if required. The initial reaction on virgin carbon has been theorized to be:



where C^* and CO^* indicate activated carbon and a carbon surface oxide, respectively.

After an initial acclimation period, the monochloramine is oxidized to nitrogen gas according to the reaction.



Although this approach appears feasible, additional extensive pilot work would be necessary prior to implementation. At the pilot test site, the cooling tower blowdown is mixed with other plant wastes, settled, and passed through a lagoon system prior to discharge into a stream. Because of this long detention time in addition to plant effluent blending, it is reasonable to assume that dechlorination facilities would not be required, thereby making the monochloramine formation process more attractive. However, for purposes of design in this report, it was assumed that the raw wastewater will require treatment to breakpoint chlorination. The pilot data indicated that a chlorine to ammonia ratio of about 10:1 would be required to consistently obtain breakpoint.

Chemical usage was monitored throughout the study and the volumes used for each run are shown in Appendix Table B-2. An average of 2 ℓ /1,000 treated (2 gal./1,000 gal. treated) of 18 percent sodium hydroxide was required to elevate the raw wastewater pH to 11.5. Neutralization was accomplished by adding 5 ℓ /1,000 liters treated (5 gal./1,000 gal. treated) of a 1 percent hydrochloric acid solution (independent of the ferric chloride added). The use of more concentrated chemicals would be expected in a full-scale plant.

Alkaline Chlorination of Reverse Osmosis Brine

Analytical data obtained from the alkaline chlorination of reverse osmosis brine are presented in Appendix Table B-3. For this treatment train, the blast furnace scrubber blowdown was chemically clarified, filtered, and passed through a spiral wound reverse osmosis unit. The resulting product water met BAT requirements, while the concentrated brine was treated by alkaline chlorination.

Table 10, columns 1 and 3 show the BAT parameters and the effluent brine concentrations allowable to meet limitations. The numbers in column 3 were calculated from EPA guidelines and blast furnace production during the study.

TABLE 10. BAT LIMITS FOR DISCHARGE OF REVERSE OSMOSIS PRODUCT WATER AND BRINE

Parameter	Concentration in mg/l		
	Allowable effluent concentrations to meet BAT limitations	Reverse ^a osmosis product water	Allowable reverse osmosis brine concentrations assuming discharge of product water
TSS	27	3	99
CN-A	0.271	0.01	1.054
Phenol	0.543	0.054	2.01
NH ₃	10.85	9	16.40
S	0.334	0.2	0.736
F	21.71	0.40	85.64
pH	6.0-9.0		

a. Projected water quality from pilot study.

An in-depth discussion of how these numbers were obtained has been previously presented in the Section 7 Introduction. Column 2 of Table 10 shows the projected reverse osmosis product water quality. BAT limitations were met for all parameters. If a 75 percent recovery is assumed, column 4 shows the pollutant concentrations that would be allowed to be discharged in the

brine, providing the product water would also be discharged instead of reused. If it is assumed that the reverse osmosis brine is discharged and the product water is reused in the plant, the allowable BAT pollutant concentrations are shown below (Table 11):

TABLE 11. ALLOWABLE BAT CONCENTRATIONS
FOR REVERSE OSMOSIS BRINE

Parameter	Allowable RO brine concentration mg/l
TSS	108
CN-A	1.084
Phenol	2.172
NH ₃	43.4
S	1.336
F	86.84
pH	6.0-9.0

Reuse of the reverse osmosis product water would be probable because of its high quality.

Brine treatment by alkaline chlorination consisted of chlorine addition at elevated pH, neutralization, and settling. Polymer was also added to enhance the settling rate.

Appendix Table B-4 presents analytical and operational data for each run conducted. A summary of Tables B-3 and B-4 is presented below in Table 12.

Influent ammonia was somewhat variable and dependent on fluctuations in the blowdown ammonia concentration. The average reverse osmosis brine feed ammonia concentration to the alkaline chlorination system was 93 mg/l with a range from 73 mg/l to 123 mg/l. Cyanide-A concentrations varied from <0.01 mg/l to 0.08 mg/l and averaged <0.03 mg/l.

Effluent pollutant concentrations were dependent on the chlorine to ammonia ratio used during the run. Typically, ammonia concentrations decreased with increasing Cl₂:NH₃ ratios. Effluent ammonia for all nine runs averaged 9.7 mg/l while the average cyanide-A in the effluent was <0.02 mg/l. A graph of influent versus effluent ammonia concentration is shown in Figure 35.

Table 13 has been prepared to illustrate the effect that the chlorine to ammonia ratio had on effluent concentrations of ammonia and cyanide-A. The table shows that ammonia remains high until the chlorine to ammonia ratio exceeded about 8:1. Breakpoint is reached at a Cl₂:NH₃ ratio of approximately

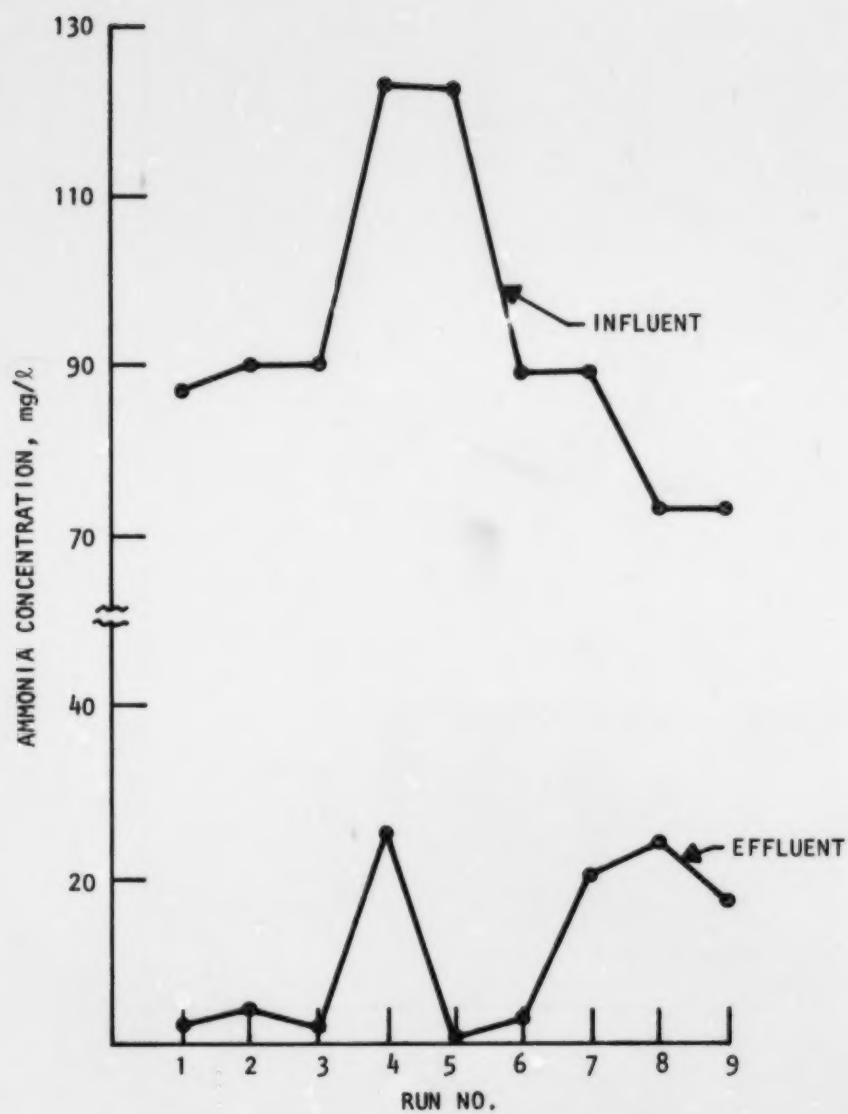


Figure 35. Influent vs. effluent ammonia concentration
alkaline chlorination - reverse osmosis brine.

TABLE 12. ALKALINE CHLORINATION OF REVERSE OSMOSIS BRINE -
WASTEWATER QUALITY DURING TESTING

	pH	SS, mg/l	NH ₃ , mg/l	CN-A, mg/l	Fluoride, mg/l	Phenol, mg/l	Sulfide, mg/l
Influent							
average	-	-	93	<0.03	18.7	0.089	<0.2
maximum	7.35	-	123	0.08	35.2	0.659	<0.2
minimum	6.1	-	73	<0.01	11.4	0.011	<0.2
Effluent							
average	-	44	9.7	<0.02	16.3	0.007	<0.2
maximum	7.85	88	25	0.03	23.5	0.015	<0.2
minimum	5.8	1	<0.05	<0.01	7.03	0.002	<0.2

TABLE 13. REVERSE OSMOSIS BRINE
ALKALINE CHLORINATION DATA

Run No.	Flow gpm	Cl ₂ :NH ₃ ratio	Ammonia, mg/l		Cyanide-A, mg/l	
			Inf.	Eff.	Inf.	Eff.
1	1.0	8.5:1	87	1.8	-	<0.01
2	2.5	8.5:1	90	4.2	<0.01	<0.01
3	2.5	6.4:1	90	2.4	<0.01	-
4	2.5	3.1:1	123	25	<0.01	0.02
5	5.0	9.6:1	123	<0.05	<0.01	<0.01
6	2.5	14.4:1	89	3.25	0.08	<0.01
7	5.0	7.25:1	89	20.5*	0.08	-
8	5.0	7.4:1	73	23.8	<0.01	0.03
9	6.0	7.0:1	73	16.8	<0.01	<0.01

* Rapid mix tank effluent.
gpm x 3.785 = l/min

10:1. Similar results were also found for alkaline chlorination of the raw wastewater. The data shown in Table 13 are plotted graphically in Figure 36 to illustrate the decrease in ammonia resulting from increased Cl₂:NH₃ ratio. Flow rates of 3.785 l/min (1 gpm), 9.46 l/min (2.5 gpm), 18.9 l/min (5.0 gpm) and 22.7 l/min (6.0 gpm) were investigated.

Some of the trends observed in alkaline chlorination of the raw wastewater were similarly observed for treatment of the reverse osmosis brine. For

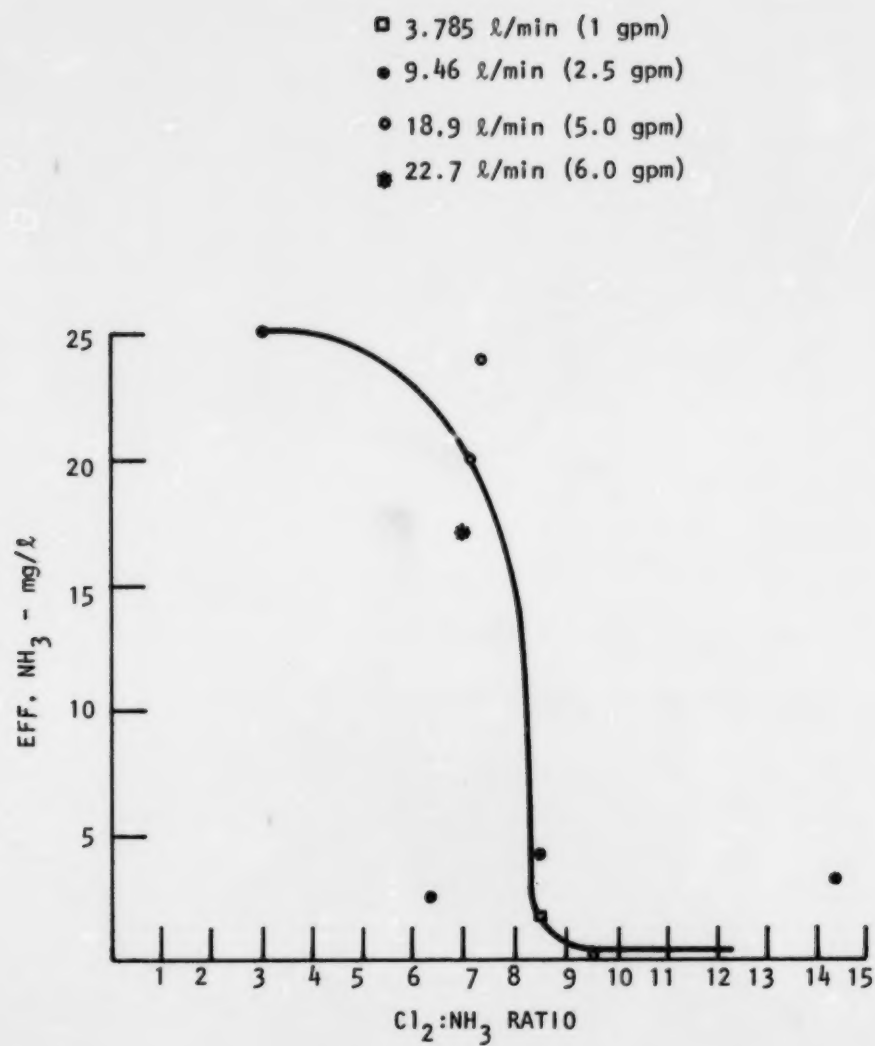


Figure 36. Effluent NH_3 vs. $\text{Cl}_2:\text{NH}_3$ ratio alkaline chlorination - reverse osmosis brine.

example, brine suspended solids into the alkaline chlorination system were initially low, but increased significantly in the first reaction chamber (alkaline conditions). As the waste was neutralized, suspended solids were reduced. The addition of polymer prior to clarification enhanced the solids settling rate to allow for effective and consistent suspended solids removal prior to dechlorination by activated carbon. For the nine runs performed, influent dissolved solids averaged 5,754 mg/l while the average effluent was 7,616 mg/l. This represents a TDS increase of 24 percent resulting from alkaline chlorination treatment of the brine. Chlorination also effectively removed influent concentrations of phenol (99 percent removal) and Total Kjeldahl Nitrogen (TKN) (89 percent removal).

Chlorine residual (measured as total chlorine) was highest in the first reaction chamber, and decreased as the waste stream passed through the system. Residual concentrations ranged from a low of 6.2 mg/l to a high of 283.6 mg/l and were dependent on the chlorine to ammonia ratio selected. All of the runs investigating alkaline chlorination of reverse osmosis brine met BAT limitations for all parameters of concern. Chlorine to ammonia ratios were varied from 3.1:1 to 14.4:1. This suggests that it is not necessary to go to breakpoint chlorination to meet BAT requirements. Similar results were also found for alkaline chlorination of the raw wastewater. Although several studies have investigated ammonia removal by monochloramine formation followed by dechlorination on activated carbon (see discussion on alkaline chlorination of raw wastewater), the results of these studies appear inconclusive. Much additional pilot test work would be required prior to full-scale application of this system. The blast furnace scrubber cooling tower blowdown from the research site is blended with other plant wastes, clarified and passed through a lagoon prior to discharge. Therefore, the possibility exists that dechlorination facilities would not be required for this particular plant. However, for the purpose of developing uniform cost information, the full-scale reverse osmosis brine alkaline chlorination system is designed for breakpoint chlorination (Cl:N ratio of 10:1). It was also assumed that dechlorination facilities would be optional, depending on local discharge limitations for chlorine.

Chemical usage for each run was measured for caustic, acid, and sodium hypochlorite. Sodium hydroxide usage averaged 5 l/1,000 liters treated (5 gal./1,000 gal. treated) using an 18 percent solution (215.5 gm/liters). Hydrochloric acid, added as a 1 percent solution (11 gm/liters) was fed to the system for neutralization at an average rate of 11.9 l/1,000 liters treated (11.9 gal./1,000 gal. treated). For a full-scale system, the use of more concentrated chemicals would be expected.

System Descriptions, Costs and Space Requirements

System Description--

The data generated in the pilot scale study were evaluated for scale up to full-size installations. Several assumptions were made in preparing the cost estimates and these have been listed previously in the Section 7

"Description of Cost Assumptions". Additional assumptions are listed below.

ASSUMPTIONS USED TO DEVELOP COST AND DESIGN FIGURES
FOR ALKALINE CHLORINATION SYSTEM AT THE BLAST FURNACE RESEARCH SITE

-
-
1. Chemical Costs (Bulk)
 - Caustic Soda (50%): \$1.46/3.785 liters (\$1.46/gal.)
 - Sulfuric Acid (66° Be): \$0.366/3.785 liters (\$0.366/gal.)
 - Ferric Chloride: Costs included in clarification design
 - Sodium Hypochlorite: \$0.55/3.785 liters (\$0.55/gal.)
 2. Chemical Costs (Bulk)
 - Chlorine Gas: \$0.33/kg (\$0.15/lb) - 237 m³/day (0.0625 mgd)
 - 2,839 m³/day (0.75 mgd),
ton cylinders
 - \$0.20/kg (\$0.09/lb) - 6,578 m³/day (1.5 mgd),
tank cars
 3. Peak NH₃ concentration = 2 x average NH₃ concentration.
 4. Design Ammonia: Raw waste - 33 mg/l; RO Brine - 93 mg/l.
-
-

Flows of 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd) and 5,678 m³/day (1.5 mgd) were used to develop costs for alkaline chlorination of the raw wastewater. For treatment of the reverse osmosis brine, design flows of 237 m³/day (0.0625 mgd), 710 m³/day (0.1875 mgd), and 1,419 m³/day (0.375 mgd) were selected. Tables 14 and 15 summarize the design basis used for the cost estimates. A design chlorine to ammonia ratio of 10:1 was selected for treatment of both the raw wastewater and the reverse osmosis brine. Both systems are designed to remove ammonia, cyanide and phenol to below BAT limits. In the full-scale system, the incoming wastewater pH would be elevated using caustic soda. Chlorine would be added and the contents mixed in-line, and fed to the cyanide treatment tank. Constant ORP and pH would be maintained by controllers and a tank mixer. After a design detention time of 20 minutes, the effluent would be pH adjusted and again blended with chlorine for ammonia removal. The contents would be mixed in-line prior to entering the ammonia treatment tank. The design detention time in this tank is 1 minute and the system is equipped with pH and ORP controllers and a mixer to maintain constant conditions. The effluent from this system would be chemically treated, clarified and dechlorinated on activated carbon (if required prior to discharge). Alarm systems for high concentrations of cyanogen chloride and nitrogen trichloride would also be provided. A schematic of the full-scale continuous alkaline chlorination treatment system is shown in Figure 37.

TABLE 14. ALKALINE CHLORINATION DESIGN DATA

Type of Waste	Raw wastewater		
Design flow, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Contact Time, min.			
cyanide removal tank	20	20	20
ammonia removal tank	1	1	1
Reaction Tank Volume, L (gal.)			
cyanide removal tank	15,140 (4,000)	43,528 (11,500)	86,109 (22,750)
ammonia removal tank	852 (225)	2,460 (650)	4,542 (1,200)
Mixing Requirements (Mechanical)			
cyanide removal tank [min. G, m/sec/m, (ft/sec/ft)]	900 (900)	900 (900)	900 (900)
ammonia removal tank [min. G, m/sec/m, (ft/sec/ft)]	900 (900)	900 (900)	900 (900)
Design Cl:N Ratio	10:1	10:1	10:1
Chemical Requirements, L/day (gal./day)			
Option #1 - Sodium Hypochlorite			
Caustic soda (50%)	568 (150)	1,703 (450)	3,217 (850)
H ₂ SO ₄ (66° Be)	38 (10)	114 (30)	227 (60)
NaOCl (15%)	2,082 (550)	6,435 (1,700)	12,680 (3,350)
Option #2 - Chlorine Gas			
Caustic soda (50%)	946 (250)	2,650 (700)	5,299 (1,400)
H ₂ SO ₄ (66° Be)	--	--	--
Chlorine gas, kg/day (lb/day)	318 (700)	953 (2,100)	1,907 (4,200)
Chemical Stock Tank Design Storage			
Caustic soda	30 day	30 day	30 day
H ₂ SO ₄	30 day	30 day	30 day
NaOCl	30 day	30 day	30 day

TABLE 15. ALKALINE CHLORINATION DESIGN DATA

Type of waste	Reverse osmosis brine		
	237 (0.0625)	710 (0.1875)	1,419 (0.375)
Design flow, m ³ /day (mgd)			
Contact Time, min.			
cyanide removal tank	20	20	20
ammonia removal tank	1	1	1
Reaction Tank Volume, l (gal.)			
cyanide removal tank	3,785 (1,000)	11,355 (3,000)	22,710 (6,000)
ammonia removal tank	227 (60)	757 (200)	1,230 (325)
Mixing Requirements (Mechanical)			
cyanide removal tank [min. G, m/sec/m, (ft/sec/ft)]	900 (900)	900 (900)	900 (900)
ammonia removal tank [min. G, m/sec/m, (ft/sec/ft)]	900 (900)	900 (900)	900 (900)
Design Cl:N Ratio	10:1	10:1	10:1
Chemical Requirements, l/day (gal./day)			
Option #1 - Sodium Hypochlorite			
Caustic soda, (50%)	379 (100)	1,022 (270)	2,082 (550)
H ₂ SO ₄ (66° Be)	38 (10)	76 (20)	151 (40)
NaOCl (15%)	1,514 (400)	4,542 (1,200)	8,706 (2,300)
Option #2 - Chlorine Gas			
Caustic soda, (50%)	568 (150)	1,741 (460)	3,407 (900)
H ₂ SO ₄ (66° Be)	--	--	--
Chlorine gas, kg/day (lb/day)	227 (500)	681 (1,500)	1,317 (2,900)
Chemical Stock Tank Design Storage			
Caustic soda	30 day	30 day	30 day
H ₂ SO ₄	30 day	30 day	30 day
NaOCl	30 day	30 day	30 day

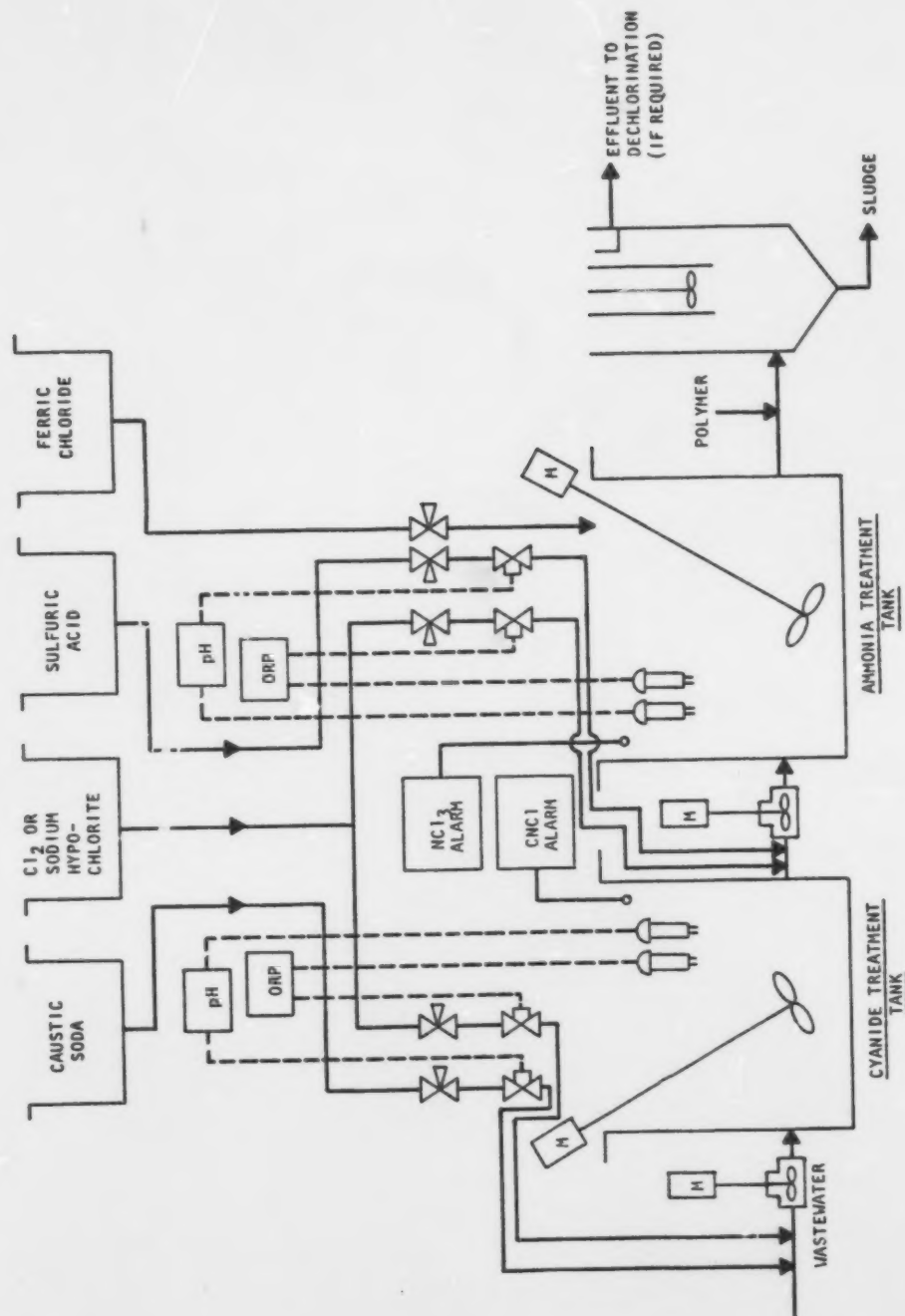


Figure 37. Schematic for full-scale continuous treatment of cyanide and ammonia.

Cost and Space Requirements--

Using the design information developed in Tables 14 and 15, alkaline chlorination treatment costs were computed for each of the design flows. Costs for the system components were based on vendor quotations. The primary installed equipment costs include:

1. Tankage.
2. Mixers.
3. pH and ORP Controllers.
4. Chemical Stock Tanks.
5. Associated Pumps, Valves, Piping and Controls.
6. Cyanogen Chloride and Nitrogen Trichloride Alarm Systems.
7. Chlorinators.

Tables 16 and 17 summarize the cost data for alkaline chlorination. The cost information in Table 16 assumes sodium hypochlorite usage while Table 17 reflects chlorine gas usage. The costs are shown for major equipment components and primary operating costs. In addition, the tables provide the expected total yearly cost for treatment in dollars and \$/3,785 liters (\$/1,000 gal.). Also shown in Table 16 and 17 are the land requirements for each of the system designs. Sufficient space has been allowed between reaction tanks to allow for aisle ways and the performance of routine maintenance.

Comparison of the two cost tables indicates that the capital cost for the two systems are similar but that the chlorine gas system has significantly lower operating cost. Therefore, chlorine gas usage would be the most economical method of treatment for all designs considered.

RESULTS OF THE CHEMICAL CLARIFICATION TREATMENT STUDY

Introduction

The purpose of chemical clarification in this study was to provide pretreatment and post-treatment for the principal technologies investigated, namely alkaline chlorination, ozonation, and reverse osmosis. A 1.8 m (6.0 ft) diameter clarifier was used to pretreat the wastewater, while a 1.3 m (4.25 ft) diameter clarifier was used for post-treatment. In both cases, the clarifiers were primarily used to remove particulates from the wastewater. The hydraulic requirements of the principal technologies listed above were controlling factors in operating the flocculator-clarifier.

Pretreatment Clarification

The 1.8 m (6.0 ft) diameter clarifier provided pretreatment for two treatment trains. The first treatment train consisted of alkaline chlorination of the raw wastewater followed by clarification and activated carbon. The second train consisted of chemical addition, clarification, filtration, and reverse osmosis (RO) treatment of the raw wastewater. A summary of the pretreatment

TABLE 16. ALKALINE CHLORINATION COST SUMMARY
(ASSUMING NaOCl FEED)

Type of Waste Design Flow Rate m ³ /day (mgd)	Raw Wastewater ^a 2,839 (0.75)	5,678 (1.5)	237 (0.0625)	710 (0.1875)	RO Brine 1,419 (0.375)
Capital Cost^b					
Pumps	\$ 5,000	\$ 6,000	\$ 5,000	\$ 5,000	\$ 6,000
Tankage	18,000	58,000	14,000	26,000	43,000
Controlers (pH & ORP)	15,000	15,000	15,000	15,000	15,000
Mixers	11,000	18,000	7,000	10,000	11,000
Electrical Panel	5,000	6,000	5,000	5,000	6,000
Alarm System	10,000	10,000	10,000	10,000	10,000
Delivered Equipment Total	\$ 64,000	\$ 114,000	\$ 56,000	\$ 71,000	\$ 91,000
Equipment Installation	26,000	46,000	22,000	28,000	36,000
Installed Equipment Total	\$ 90,000	\$ 160,000	\$ 78,000	\$ 99,000	\$ 127,000
Process Piping	27,000	48,000	23,000	30,000	38,000
Instrumentation	5,000	8,000	4,000	5,000	6,000
Capital Equipment Total	\$ 122,000	\$ 216,000	\$ 105,000	\$ 134,000	\$ 171,000
Facilities & Fees	262,000	465,000	226,000	288,000	368,000
Capital Cost Total	\$ 384,000	\$ 681,000	\$ 331,000	\$ 422,000	\$ 539,000
Operating Cost					
Electrical	\$ 1,000	\$ 4,000	\$ 1,000	\$ 1,000	\$ 1,000
Operator	54,000	54,000	54,000	54,000	54,000
Maintenance Labor & Materials	6,000	11,000	5,000	7,000	9,000
Chemicals	194,000	1,133,000	134,000	387,000	760,000
Yearly Operating Subtotal	\$ 255,000	\$ 1,202,000	\$ 191,000	\$ 449,000	\$ 824,000
Capital Amortization (20 yr life @ 6.875%)	36,000	64,000	31,000	39,000	50,000
Total Yearly Cost (\$/yr)	\$ 291,000	\$ 1,266,000	\$ 222,000	\$ 488,000	\$ 874,000
\$/3,785 & (\$/1,000 gal.)	3.19	2.56	9.73	7.13	6.39
Land Requirement m ² (ft ²)	186 (2,000)	362 (3,900)	149 (1,600)	230 (2,475)	322 (3,500)

a. If uniform flow to the alkaline chlorination system cannot be maintained, an equalization basin may be required.

There has been no cost allowance for equalization in the above designs.

b. See Appendix Tables A-4 and A-5 for basis of estimate.

TABLE 17. ALKALINE CHLORINATION COST SUMMARY
(ASSUMING Cl₂ FEED)

Type of Waste	Raw Wastewater ^a				RO Brine	
Design Flow Rate m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)	237 (0.0625)	710 (0.1875)	1,419 (0.375)
Capital Cost^b						
Pumps	\$ 2,000	\$ 2,000	\$ 2,000	\$ 2,000	\$ 2,000	\$ 2,000
Tankage	8,000	17,000	26,000	5,000	10,000	16,000
Controllers (pH & ORP)	15,000	15,000	15,000	15,000	15,000	15,000
Mixers	11,000	14,000	18,000	7,000	10,000	11,000
Electrical Panel	5,000	6,000	7,000	5,000	5,000	6,000
Alarm System	10,000	10,000	10,000	10,000	10,000	10,000
Chlorination	20,000	25,000	25,000	20,000	25,000	25,000
Delivered Equipment Total	\$ 71,000	\$ 89,000	\$103,000	\$ 64,000	\$ 77,000	\$ 85,000
Equipment Installation	28,000	36,000	41,000	26,000	31,000	34,000
Installed Equipment Total	\$ 99,000	\$125,000	\$144,000	\$ 90,000	\$108,000	\$119,000
Process Piping	30,000	38,000	43,000	27,000	32,000	36,000
Instrumentation	5,000	6,000	7,000	5,000	5,000	6,000
Capital Equipment Total	\$134,000	\$169,000	\$194,000	\$122,000	\$145,000	\$161,000
Facilities & Fees	287,000	364,000	417,000	262,000	312,000	346,000
Capital Cost Total	\$421,000	\$533,000	\$611,000	\$384,000	\$457,000	\$507,000
Operating Cost						
Electrical	\$ 1,000	\$ 2,000	\$ 4,000	\$ 1,000	\$ 1,000	\$ 1,000
Operator	54,000	54,000	54,000	54,000	54,000	54,000
Maintenance Labor & Materials	7,000	8,000	10,000	6,000	7,000	8,000
Chemicals	172,000	488,000	884,000	107,000	327,000	480,000
Yearly Operating Subtotal	\$234,000	\$552,000	\$952,000	\$168,000	\$389,000	\$543,000
Capital Amortization						
(20 yr life @ 6.875%)	39,000	50,000	57,000	26,000	43,000	48,000
Total Yearly Cost (\$/yr)	\$273,000	\$602,000	\$1,009,000	\$204,000	\$432,000	\$591,000
\$/3,785 L (\$/1,000 gal.)	2.99	2.20	1.84	8.94	6.31	4.32
Land Requirement m ² (ft ²)	163 (1,750)	204 (2,200)	279 (3,000)	139 (1,500)	153 (1,650)	179 (1,925)

a. If uniform flow to the alkaline chlorination system cannot be maintained, an equalization basin may be required.
b. See Appendix Tables A-4 and A-5 for basis of estimate.

clarifier operating parameters in conjunction with influent and effluent quality is given in Appendix Table C-1. The data in Table C-1 are summarized by overflow rates. Turbidity and pH measurements were taken on-site while the values reported for SS, NH_3 , CN^- , phenol, F^- , and sulfide were obtained from 24 hour composite samples. Additional test parameters of interest monitored during the study can be found in the Appendix Table A-1 Data Summary.

Alkaline Chlorinated Raw Wastewater--

Alkaline chlorinated raw wastewater was chemically treated and clarified from January 29, 1979 to February 22, 1979.

Clarification was included in this treatment train to remove suspended solids present in the alkaline chlorinated raw wastewater prior to activated carbon dechlorination. Ferric chloride (FeCl_3) and Betz 1100 polyelectrolyte were used to enhance clarification. Dosages typically were 100 mg/l FeCl_3 and 0.5 mg/l polymer. Bench clarification tests during the study determined that these dosages produced optimum effluent quality.

Table 18 presents the suspended solids removals obtained during the study. Flow rates of $6.1 \text{ m}^3/\text{m}^2/\text{day}$ (150 gpd/ft²), $10.2 \text{ m}^3/\text{m}^2/\text{day}$ (250 gpd/ft²), and $16.3 \text{ m}^3/\text{m}^2/\text{day}$ (400 gpd/ft²) were investigated. The suspended solids removal averaged 93 percent with a range of 88-98 percent. As can be seen in Table 18, average percent removals of the three overflow rates are essentially equal. The average settling rate was determined to be 0.2 m/min. (0.7 ft/min.).

Sludge production during this portion of the study averaged 40 ℓ /1,000 liters (40 gal./1,000 gal.) at an overflow rate of $10.2 \text{ m}^3/\text{m}^2/\text{day}$ (250 gpd/ft²) and 30 ℓ /1,000 liters (30 gal./1,000 gal.) at an overflow rate of $16.3 \text{ m}^3/\text{m}^2/\text{day}$ (400 gpd/ft²). Table 19 presents the suspended and volatile suspended solids concentrations of grab samples of the sludge taken during the study. At an overflow rate of $10.2 \text{ m}^3/\text{m}^2/\text{day}$ (250 gpd/ft²), the sludge suspended solids concentration averaged 3,900 mg/l and 2,300 mg/l at an overflow rate of $16.3 \text{ m}^3/\text{m}^2/\text{day}$ (400 gpd/ft²). A thickening test was performed on the sludge during the study. The sludge was settled for 60 minutes in a one liter cylinder. The resulting thickened sludge concentration was 7,100 mg/l and the decant solids was 3 mg/l. The influent sludge concentration was about 3,000 mg/l. A settling rate of 0.28 cm/min. (0.009 ft/min.) was determined in the thickening test.

Raw Wastewater--

Blast furnace gas scrubber system blowdown was chemically treated and clarified from February 29, 1979 to March 26, 1979. Chemical treatment and clarification, along with filtration was required as pretreatment for the downstream reverse osmosis unit. The objective of this portion of the study was to reduce the suspended solids and turbidity of the influent raw wastewater. Chemical treatment consisted of the addition of alum and Betz 1100 polymer. Alum dosages ranged from 50 mg/l to 150 mg/l, while polymer was added at 0.5 mg/l. These dosages were adjusted when necessary to produce

TABLE 18. SUSPENDED SOLIDS REMOVAL OF ALKALINE CHLORINATED WASTEWATER

Date	Overflow rate, gpd/ft ²	Suspended solids, mg/l		Percent removal
		Inf.	Eff.	
2-11-79	150	200	25	88
2-12-79	150	192	11	94
2-13-79	150	1320	33	98
2-6-79	250	137	16	88
2-7-79	250	220	16	93
2-8-79	250	264	16	94
2-14-79	250	1455	35	98
2-15-79	400	2070	36	98
2-19-79	400	320	29	91
2-20-79	400	290	25	91
2-21-79	400	450	37	92
2-22-79	400	490	36	93

Note: gpd/ft² x 0.0407 = m³/m²/day

TABLE 19. ALKALINE CHLORINATED WASTEWATER SLUDGE SUSPENDED SOLIDS CONCENTRATIONS

Date	Overflow rate, gpd/ft ²	SS, mg/l	VSS, mg/l	Sludge wasting rate ^a , gph
2-1-79	250	39,900	2,200	9
2-4-79	250	4,090	425	18
2-6-79	250	2,470	250	-
2-7-79	250	1,540	205	11
2-10-79	250	6,340	3,110	10
2-11-79	250	4,530	610	10
2-12-79	250	4,010	650	13
2-13-79	250	4,140	910	11
2-14-79	250	14,725	2,740	16
2-15-79	400	775	135	13
2-16-79	400	5,550	900	20
2-19-79	400	3,430	460	14
2-20-79	400	1,265	235	64
2-21-79	400	2,505	330	70
2-22-79	400	480	135	63

a. Instantaneous values, not intended to represent average daily rates.
gpd/ft² x 0.0407 = m³/m²/day gph x 3.785 = l/hr

optimum effluent quality, as determined by bench settling tests performed on-site. Alum was used as a coagulant, instead of FeCl_3 to eliminate any potential reverse osmosis iron fouling associated with the use of ferric chloride. A summary of the suspended solids and turbidity removals obtained during the study is presented in Table 20. Appendix Table A-2 presents all data obtained from chemical treatment and clarification of raw wastewater.

Sludge production was not determined during the test program. Previous bench scale testing indicated that 8 liters of sludge would be produced for every 1,000 liters of raw wastewater (8 gal./1,000 gal.).

TABLE 20. SUSPENDED SOLIDS AND TURBIDITY REMOVALS
FOR CHEMICAL CLARIFICATION OF RAW WASTEWATER

	Suspended solids, mg/l			Turbidity, FTU		
	Inf.	Eff.	Percent removal	Inf.	Eff.	Percent removal
Maximum	5734	484	92	53	40	24
Minimum	31	16	48	23	2.0	91
Average	154	26	83	38	13	66

Discussion--

Chemical treatment and clarification of the alkaline chlorinated raw wastewater was necessary to remove suspended solids present in the wastewater to reduce the solids loadings on the activated carbon adsorption system. High solids loading to an activated carbon system would increase the plugging of the carbon columns with entrained solids and increase the frequency of backwashes necessary to remove the entrapped solids. The low effluent suspended solids indicated that clarification was providing adequate pretreatment for the activated carbon.

Chemical treatment and clarification of the raw wastewater was necessary to reduce the suspended solids concentration and turbidity of the wastewater prior to filtration. Reverse osmosis requires feedwater with low suspended solids and turbidity concentrations to prevent fouling of the membranes. Operation of the reverse osmosis pilot unit over the study period indicated no significant fouling problems. This confirms that the pretreatment consisting of chemical addition, clarification, and filtration was adequate for operation of a reverse osmosis unit.

Post-treatment Clarification

The 1.3 m (4.25 ft) diameter clarifier was used for final treatment for three of the process trains studied. The first treatment train consisted of filtration and ozonation of the raw wastewater followed by clarification.

The second and third treatment trains consisted of settling the alkaline chlorinated or ozonated reverse osmosis brine.

Ozonated Filter Effluent--

Ozonated filter effluent was clarified from January 10, 1979 to January 20, 1979. Clarification was included in this treatment train in order to remove the suspended solids present in the ozonated effluent. Betz 1100 polyelectrolyte was used as a coagulant aide, typically at a dose of 0.5 mg/l.

A summary of the performance data obtained for clarification of ozonated filter effluent is presented in Table 21. Suspended solids removal averaged 97 percent at an overflow rate of $5.1 \text{ m}^3/\text{m}^2/\text{day}$ ($125 \text{ gpd}/\text{ft}^2$) and a 9.6 hour retention time. A $10.2 \text{ m}^3/\text{m}^2/\text{day}$ ($250 \text{ gpd}/\text{ft}^2$) overflow rate and a 2.8 hour retention time also achieved a 97 percent suspended solids removal. The settling rate was determined to be $1.83 \text{ cm}/\text{min}$. ($0.06 \text{ ft}/\text{min}$.) on January 13, 1979.

TABLE 21. CLARIFICATION OF OZONATED FILTER EFFLUENT

Date	Overflow rate, gpd/ft^2	Suspended solids, mg/l		Percent removal
		Influent	Effluent	
1-13-79	125	45	1	98
1-15-79	125	296	3	99
1-16-79	125	196	9	95
1-19-79	150	-	8	-
1-20-79	175	-	15	-
1-17-79	250	248	7	97
1-10-79	500	195	-	-

Note: $\text{gpd}/\text{ft}^2 \times 0.0407 = \text{m}^3/\text{m}^2/\text{day}$

The average sludge production rate was $20 \text{ l}/1,000 \text{ liters}$ ($20 \text{ gal.}/1,000 \text{ gal.}$) wastewater. Table 22 gives the suspended and volatile solids data obtained during this portion of the study. A sludge thickening test, performed on January 13, 1979, showed the sludge solids concentration could be increased from $3,168 \text{ mg}/\text{l}$ to $22,290 \text{ mg}/\text{l}$ after 30 minutes of settling. The decant had a suspended solids concentration of $3 \text{ mg}/\text{l}$. The solids concentrations given in Table 22 indicate that sludge thickening took place in the clarifier during the detention times utilized.

Alkaline Chlorinated RO Brine--

Alkaline chlorinated RO brine was clarified from March 20, 1979 to March

TABLE 22. OZONATED FILTER EFFLUENT SLUDGE SOLIDS DATA

Date	Overflow rate, gpd/ft ²	Suspended solids, mg/l	Volatile suspended solids, mg/l
1-13-79	125	3,168	148
1-15-79	125	46,830	-
1-16-79	125	52,300	2,200
1-19-79	150	60,700	2,400
1-17-79	250	43,500	-
1-18-79	250	23,800	1,300
1-10-79	500	12,000	600
1-11-79	500	126,000	5,000

Note: $\text{gpd/ft}^2 \times 0.04074 = \text{m}^3/\text{m}^2/\text{day}$

26, 1979. Clarification was included in this treatment train in order to remove residual suspended solids following alkaline chlorination and therefore decrease the solids loading on the activated carbon dechlorination system. Betz 1100 anionic polymer was used as a coagulant aide, typically at a dose of 0.5 mg/l.

Table 23 presents a summary of the suspended solids removal for this portion of the test program. Suspended solids removals varied from a low of 53 percent to a maximum 97 percent. Removal appeared to be independent of the hydraulic loading.

TABLE 23. SUSPENDED SOLIDS REMOVAL OF ALKALINE CHLORINATED RO BRINE

Date	Overflow rate, gpd/ft ²	Suspended solids, mg/l		Percent removal
		Influent	Effluent	
3-20-79	100	-	88	-
3-21-79	250	143	54	62
3-23-79	250	834	76	91
3-22-79	500	90	42	53
3-24-79	500	330	-	-
3-25-79	500	881	-	-
3-26-79	600	1,366	34	97

Note: $\text{gpd/ft}^2 \times 0.04074 = \text{m}^3/\text{m}^2/\text{day}$

Sludge solids concentrations were determined twice during the test program. The solids concentration was 27,740 mg/l on March 20 and 363 mg/l on March 21. A thickening test, performed on March 24 indicated that the sludge could be thickened to a concentration of 54,200 mg/l in one hour with a supernatant solids concentration of 36 mg/l.

Ozonated RO Brine--

Batch settling tests were performed on ozonated RO brine from March 7 to March 16, 1979. The 1.3 m (4.25 ft) diameter clarifier was not utilized due to the low flow rates and subsequent impractical retention times dictated by the reverse osmosis unit's operational characteristics and the limited brine storage capacity. A 18.9 liters (5 gal.) container was used to simulate a clarifier in the batch settling tests. A 30 minute retention time was utilized. Clarification was required in this treatment train to remove the residual suspended solids produced during the ozonation process. Betz 1100 anionic polymer was used as a coagulant aide, typically at a dose of 1.0 mg/l.

The suspended solids removal for clarification of ozonated RO brine is presented in Table 24. The 98 percent removal indicates that a good settling sludge was produced. The high settling rate of 60 cm/min (2 ft/min) is also an indication of a good settling sludge.

Sludge production in the batch tests averaged 20 g/l,000 liters (20 gal./1,000 gal.). Table 25 gives the sludge solids concentrations obtained during the test program. No sludge thickening data were obtained.

Discussion--

Post-treatment clarification would be required to meet BAT discharge limitations for suspended solids. The treatment train consisting of filtration, ozonation, and clarification of the raw wastewater would be required to meet an effluent suspended solids limitation of 27.1 mg/l. As shown in Table 21, the effluent limitation is easily met at all overflow rates investigated.

Clarification following reverse osmosis brine treatment would be required to meet different suspended solids discharge limitations since the effluent guidelines are established at 0.013 kg TSS/kg product (0.013 lb/1,000 lb product). The low suspended solids in the reverse osmosis product stream allows for brine discharge of up to 99 mg/l TSS in the treated brine at 75 percent product water recovery when the RO product and treated brine are discharged together. If the treated brine is discharged separately and the product returned for use in the mill, the allowable suspended solids increases to 108 mg/l because of the 25 percent reduction in volume discharged. The clarified alkaline chlorinated brine and ozonated brine meet both suspended solids limitations.

Full-Scale Clarification Design

The results of the bench scale and pilot scale study indicated that a conventional clarification system with an overflow rate of 22 m³/m²/day

TABLE 24. OZONATED BRINE SUSPENDED SOLIDS REMOVAL

Date	Suspended solids, mg/l		Percent removal
	Influent	Effluent	
3-7-79	-	38	-
3-8-79	-	31	-
3-9-79	2,390	39	98
3-10-79	3,377	76	98
3-11-79	2,375	39	98
3-12-79	-	-	-
3-13-79	-	74	-
3-14-79	-	-	-
3-15-79	-	53	-
3-16-79	-	65	-

TABLE 25. OZONATED BRINE SLUDGE SOLIDS CONCENTRATIONS

Date	Suspended solids,	Volatile suspended solids,
	mg/l	mg/l
3-7-79	44,150	38,880
3-9-79	34,790	2,270
3-10-79	32,990	2,910
3-11-79	6,040	1,170
3-13-79	28,660	2,390
3-14-79	11,940	-
3-15-79	31,190	3,080
3-16-79	52,110	-

(530 gpd/ft²) would effectively reduce the suspended solids and turbidity for all five of the treatment trains investigated. System costs and space requirements based on vendor estimates, were prepared for each system at three typical flow rates. The costs are effective as of May, 1979. Table 26 presents the cost summary for the pretreatment systems. The cost summary for the post-treatment systems is given in Table 27.

The plant costs include chemical feed systems, sludge pumps, piping, concrete for clarifiers, excavation, valves, electrical equipment, installation, land acquisition, buildings, and engineering fees. Chemical costs are based on the dosage rates proven effective during the pilot study. Maintenance labor and materials were estimated at 5 percent of capital equipment costs. Electrical costs were based on \$0.03/KWH. The capital costs were amortized over a 20 year service life using a 6.875 percent interest rate. The manpower estimates are based on 50 percent skilled (\$18,000/year) and 50 percent unskilled (\$16,000/year) and are taken from EPA Process Design Manual No. 11 14-12-930, "Suspended Solids Removals", October 1979 (11).

RESULTS OF THE DUAL MEDIA FILTRATION STUDY

Introduction

Dual media filtration tests were performed on the blast furnace blowdown from December, 1978 through March, 1979. During December and January, filtration tests of the raw blowdown were performed. During February and March, filtration tests on blowdown chemically pretreated with alum and polymer were conducted. The filtration media used in the tests consisted of 38 cm (15 in.) of Red Flint filter sand (effective size 0.47 mm) and 43 cm (17 in.) of anthracite coal (effective size 0.94 mm).

Tables D-1 and D-3 in the Appendix present the suspended solids analysis of all 24 hour composites taken during the study. Included in these tables are information on influent and effluent quality and solids loading for each run of the dual media filter. Two identifiable types of feed were filtered as shown in the summary Tables D-2 and D-4. These included raw and chemically pretreated blowdown.

Filtration of Raw Blast Furnace Blowdown

Dual media filtration results of the raw blast furnace blowdown are summarized in Table 28. Filtration removed significant quantities of suspended solids (83 percent removal) and turbidity (75 percent removal).

Filtration of Chemically Pretreated Blast Furnace Blowdown

Table 29 shows the average, maximum and minimum influent and effluent characteristics when chemically pretreated blowdown was filtered. Overall removal of suspended solids (44 percent removal) and turbidity (66 percent removal) were lower than those observed for the raw blowdown.

TABLE 26. PRETREATMENT CHEMICAL CLARIFICATION COST SUMMARY
BLAST FURNACE SITE

	Alkaline chlorinated raw wastewater		Raw wastewater ^a	
System size, m ³ /day , (mgd)	946 (0.25)	2,839 (0.75)	946 (0.25)	2,839 (0.75)
Clarifier diameter, m(ft)	7.6(25)	12.2(40)	7.6(25)	12.2(40)
				18.3(60)
<u>Capital Costs^b</u>				
Equipment	\$ 84,000	\$142,000	\$ 84,000	\$142,000
Facilities & Fees	117,000	199,000	117,000	199,000
Total Capital Costs	\$201,000	\$341,000	\$201,000	\$341,000
				\$181,000
				252,000
				\$433,000
<u>Operating Costs</u>				
Chemical	\$ 6,000 ^c	\$ 17,000 ^c	\$ 4,000 ^d	\$ 12,000 ^d
Electrical	1,000	2,000	1,000	2,000
Manpower	10,000	17,000	10,000	17,000
Maintenance & Supplies	4,000	7,000	4,000	7,000
Subtotal	\$ 21,000	\$ 43,000	\$ 19,000	\$ 38,000
Amortization	19,000	32,000	19,000	32,000
Total Yearly Cost	\$ 40,000	\$ 75,000	\$ 38,000	\$ 70,000
				\$ 61,000
				40,000
				\$101,000
\$/3.8 m ³ (\$/1,000 gal.)	\$0.44	\$0.27	\$0.42	\$0.26
				\$0.18
<u>Space Requirements</u>				
m ²	50	130	50	130
(ft ²)	(540)	(1,400)	(540)	(1,400)
				300
				(3,100)

a. If uniform flow to the clarifiers cannot be maintained, an equalization basin may be required.
There has been no cost allowance for equalization in the above designs.

b. See Appendix Tables A-4 and A-5 for basis of estimate.

c. Cost combined for FeCl₃ and Betz 1100 polymer. Does not include shipping.
FeCl₃ (420 Be), \$109/metric ton (\$120/ton). Betz 1100 \$6.50/kg (\$2.95/lb).

d. Cost combined for alum and Betz 1100 polymer. Does not include shipping. Alum \$132/
metric ton (\$146/ton). Betz 1100 \$6.50/kg (\$2.95/lb).

TABLE 27. POST-TREATMENT CHEMICAL CLARIFICATION COST SUMMARY
BLAST FURNACE SITE

	Ozonated filter effluent		Alkaline chlorinated reverse osmosis brine		Ozonated reverse osmosis brine	
System size, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)	710 (0.1875)	1,420 (0.375)	1,420 (0.375)
Clarifier Diameter, m(ft)	7.6(25)	12.2(40)	18.3(60)	6.1(20)	9.1(30)	9.1(30)
Capital Costs ^a						
Equipment	\$ 84,000	\$142,000	\$181,000	\$ 81,000	\$107,000	\$107,000
Facilities & Fees	117,000	199,000	252,000	112,000	150,000	150,000
Total Capital Costs	\$201,000	\$341,000	\$433,000	\$193,000	\$257,000	\$257,000
Operating Costs						
Chemical ^b	\$ 1,000	\$ 4,000	\$ 7,000	\$ 2,000	\$ 4,000	\$ 4,000
Electrical	1,000	2,000	3,000	1,000	1,000	1,000
Manpower	10,000	17,000	25,000	10,000	10,000	10,000
Maintenance & Supplies	4,000	7,000	9,000	4,000	5,000	5,000
Subtotal	\$ 16,000	\$ 30,000	\$ 44,000	\$ 17,000	\$ 20,000	\$ 20,000
Amortization	19,000	32,000	40,000	18,000	24,000	24,000
Total Yearly Cost	\$ 35,000	\$ 62,000	\$ 84,000	\$ 35,000	\$ 44,000	\$ 44,000
\$/3.8 m ³ (\$/1,000 gal.)	\$0.38	\$0.23	\$0.15	\$1.23	\$0.51	\$0.32
Space Requirements						
m ² (ft ²)	50 (540)	130 (1,400)	300 (3,100)	12 (120)	30 (350)	70 (780)

a. See Appendix Tables A-4 and A-5 for basis of estimate.

b. Cost for Betz 1100 polymer \$6.50/kg (\$2.95/lb). Does not include shipping.

TABLE 28. DUAL MEDIA FILTRATION OF RAW BLAST FURNACE
BLOWDOWN-INFLUENT VS EFFLUENT QUALITY

	pH	SS, mg/l	Temp. °C	Turbidity FTU
			<u>Influent</u>	
Average	--	84	22	38
Maximum	7.8	122	24	50*
Minimum	6.6	51	18	23
			<u>Effluent</u>	
Average	--	14	22	9.5
Maximum	7.8	20	24	20
Minimum	6.6	9	18	1.1

* On isolated occasions, the darkness of the water prevented proper light penetration resulting in readings lower than actual.

TABLE 29. DUAL MEDIA FILTRATION OF CHEMICALLY
PRETREATED BLOWDOWN-INFLUENT VS EFFLUENT QUALITY

	pH	SS, mg/l	Temp. °C	Turbidity FTU
			<u>Influent</u>	
Average	--	27	25	12.2
Maximum	7.3	28	28	26
Minimum	6.7	25	21	6.5
			<u>Effluent</u>	
Average	--	15	25	4.1
Maximum	7.3	21	29	7.0
Minimum	6.7	10	21	1.2

Data Collection

In addition to overall performance data, the effect of specific operating parameters was determined for each filter run. Since suspended solids removal is the most important measure of filter performance, this parameter was monitored during each run in addition to the filtration rate, headloss and run time.

Accumulation of solids causes increasing pressure drop across the media. When sufficient solids have been deposited in the media, the headloss exceeds the design influent head and the filter must be backwashed to maintain the desired filtration rate. During any pilot study, the feed suspended solids varies from run to run and rarely equals the design conditions. Therefore, the most meaningful parameter of filter performance is the amount of solids trapped by the filter per unit area per unit headloss. This is usually reported in terms of kilograms suspended solids/m²/m headloss (pounds suspended solids/ft²/ft H.L.).

Appendix Tables D-1, D-2, D-3 and D-4 list all filter runs where complete information on filtration rate, headloss, run time, suspended solids, etc. were available. Also tabulated are the calculated solids loading capacities of each run. The tables separate runs with raw blast furnace blowdown (Tables D-1 and D-2) and those with chemically pretreated blowdown (Tables D-3 and D-4).

Discussion of Results

Raw Blast Furnace Blowdown--

The data tabulated in the Appendix were used to characterize the filter operation for design purposes. From overall analysis of the raw blast furnace blowdown it has been determined that 84 mg/l suspended solids concentration is representative of this influent. By plotting individual filter run influent against effluent suspended solids it is possible to obtain additional insight into the feed characteristics. Notice in Figure 38 that the raw blowdown had an average suspended solids removal efficiency of 83 percent.

Table 30 summarizes the average solids loading for various filtration rates. These data are presented graphically in Figure 39 to show the relationship between solids loading and filtration rate.

As seen in the table and graph, solids loadings at different flow rates varied significantly from 9.871 kg SS/m²/m H.L. (0.616 lb SS/ft²/ft H.L.) to 1.876 kg SS/m²/m H.L. (0.117 lb SS/ft²/ft H.L.).

Basis for optimum filtration rate selection--It is necessary to select a filtration rate which will result in adequate effluent quality while providing long enough cycle times to minimize backwash water usage. As shown in Figure 40 the effluent quality was dependent on filtration rate with lower removal of suspended solids as the rate was increased. Assuming an

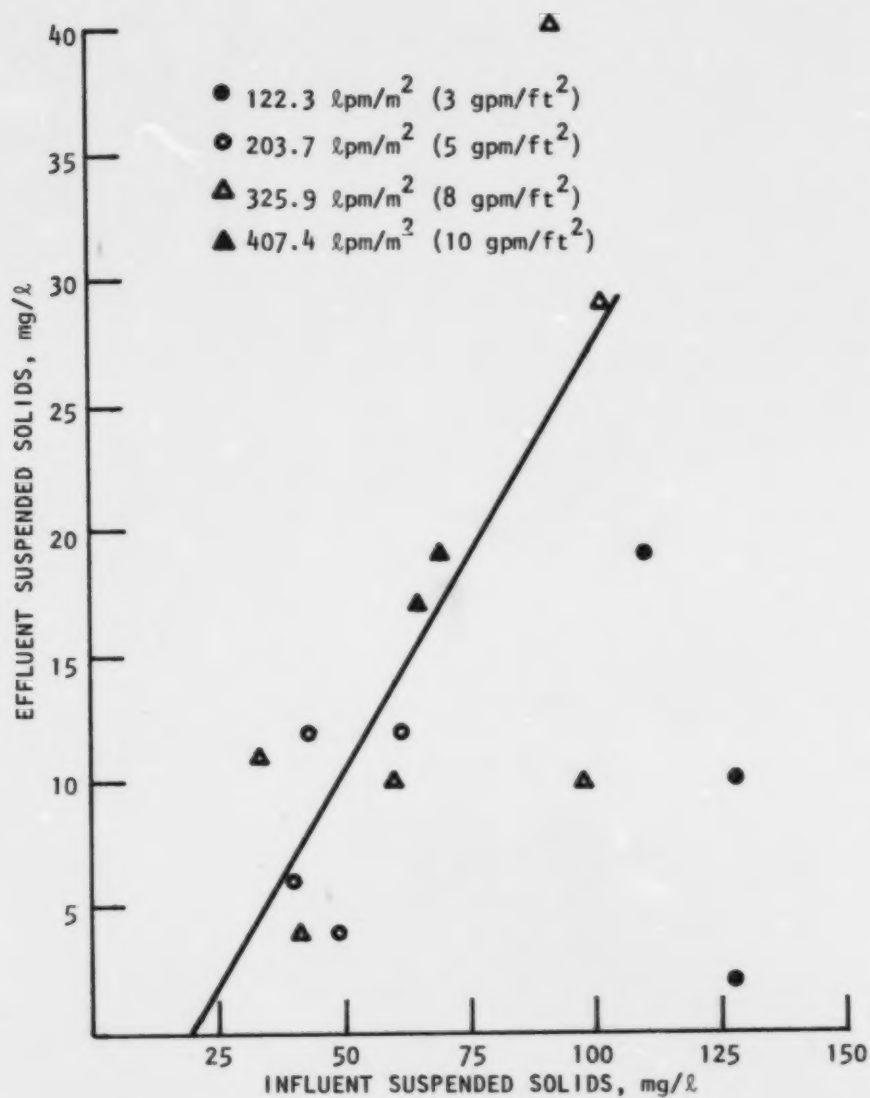


Figure 38. Dual media filter relationship of effluent to influent suspended solids for raw blast furnace blowdown.

TABLE 30. SUMMARY OF DUAL MEDIA FILTER DATA FOR
FILTRATION WITH RAW BLAST FURNACE BLOWDOWN

Filtration rate		Average solids loading	
ℓ/min/m ²	gpm/ft ²	kg SS/m ² /m H.L.	lb SS/ft ² /ft H.L.
122.2	3	6.628	0.411
203.7	5	9.871	0.616
325.9	8	3.750	0.234
407.4	10	1.876	0.117

Note: 1b SS/ft²/ft H.L. x 16 = kg SS/m²/m H.L.
gpm/ft² x 40.74 = ℓ/min/m²

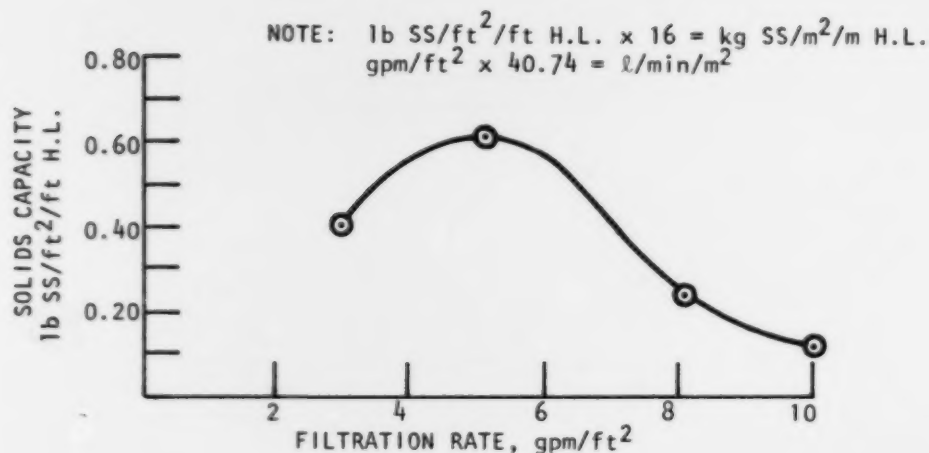


Figure 39. Dual media filter solids
capacity with raw blast furnace blowdown.

average feed suspended solids of 84 mg/l and a required effluent quality of 15 mg/l suspended solids, Figure 40 shows the maximum permissible filtration rate is 265 ℓ/min/m² (6.5 gpm/ft²).

Under normal operation, for control reasons, it is desirable to limit the frequency of backwash to once each 12 hours. Also to prevent anaerobic problems the maximum time span between backwashing is 36 hours. From

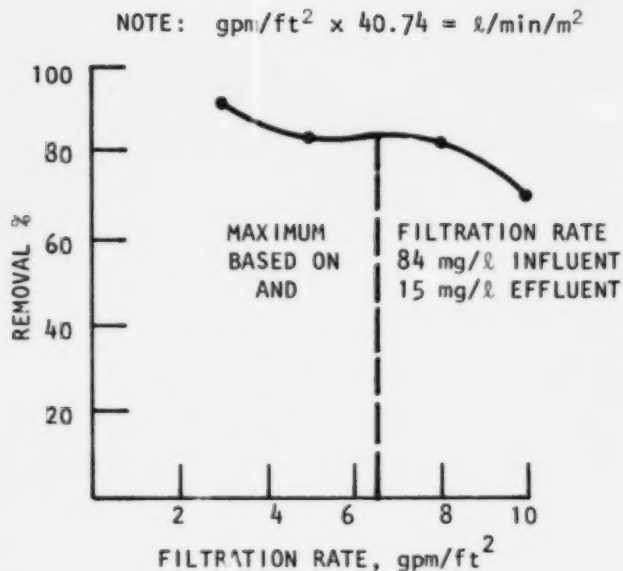


Figure 40. Effect of filtration rate on solids removal of raw blast furnace blowdown.

Figure 41, showing run time versus filtration rate, a filtration rate of 203 l/min/m^2 (5 gpm/ft^2) with 9.1 m (30 ft) of headloss would run in excess of 36 hours. For the same rate but with a terminal headloss of 10 ft the maximum duration between backwashing would be 30 hours. For design purposes, based on effluent requirements, backwash frequency and equipment design, a filtration rate of 203 l/min/m^2 (5 gpm/ft^2) was chosen for gravity and pressure filtration.

Chemically Pretreated Blast Furnace Blowdown--

An evaluation of the chemically pretreated blowdown indicates that a 27 mg/l suspended solids concentration is representative of this wastewater. By plotting individual filter run influent against effluent suspended solids, it is possible to obtain additional insight into the feed characteristics. Examining Figure 42 shows that chemically pretreated blowdown had an average suspended solids removal efficiency of 44 percent and that the filter removal efficiency can be somewhat variable. Similar results were obtained for filtration of the raw wastewater.

NOTE: $\text{ft} \times .3048 = \text{m}$
 $\text{gpm}/\text{ft}^2 \times 40.74 = \text{l}/\text{min}/\text{m}^2$

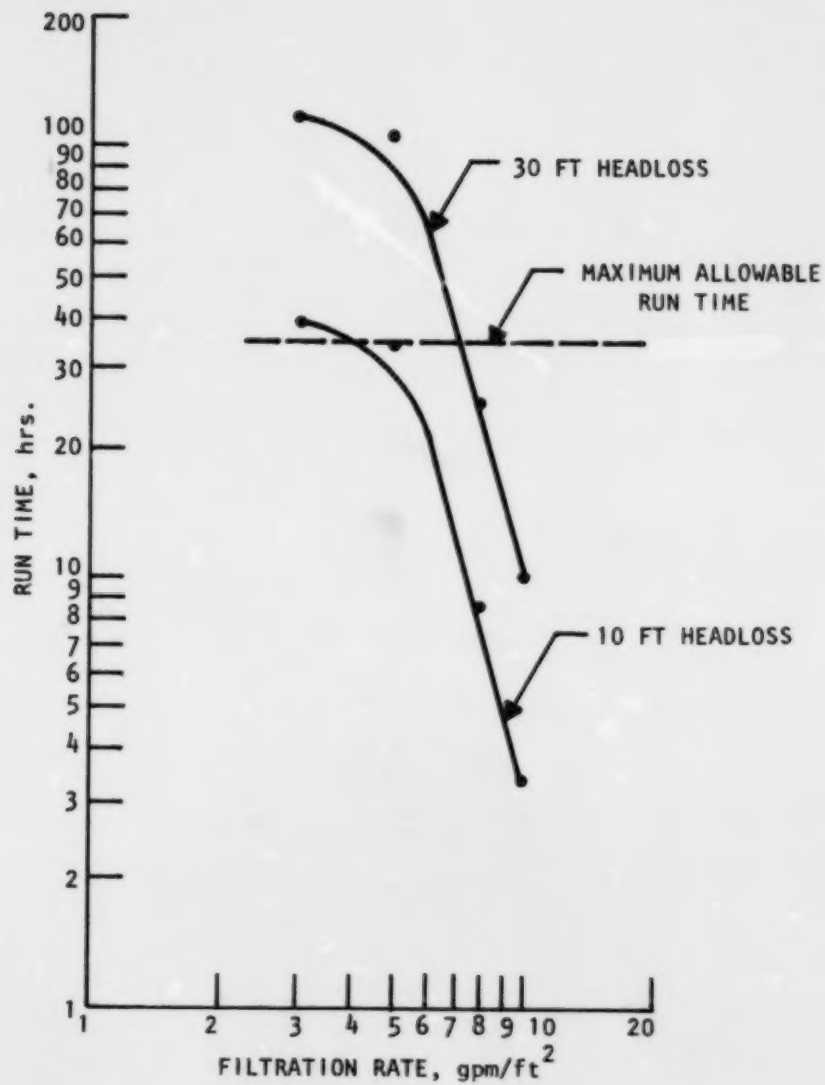


Figure 41. Dual media filter run time-raw blast furnace blowdown.

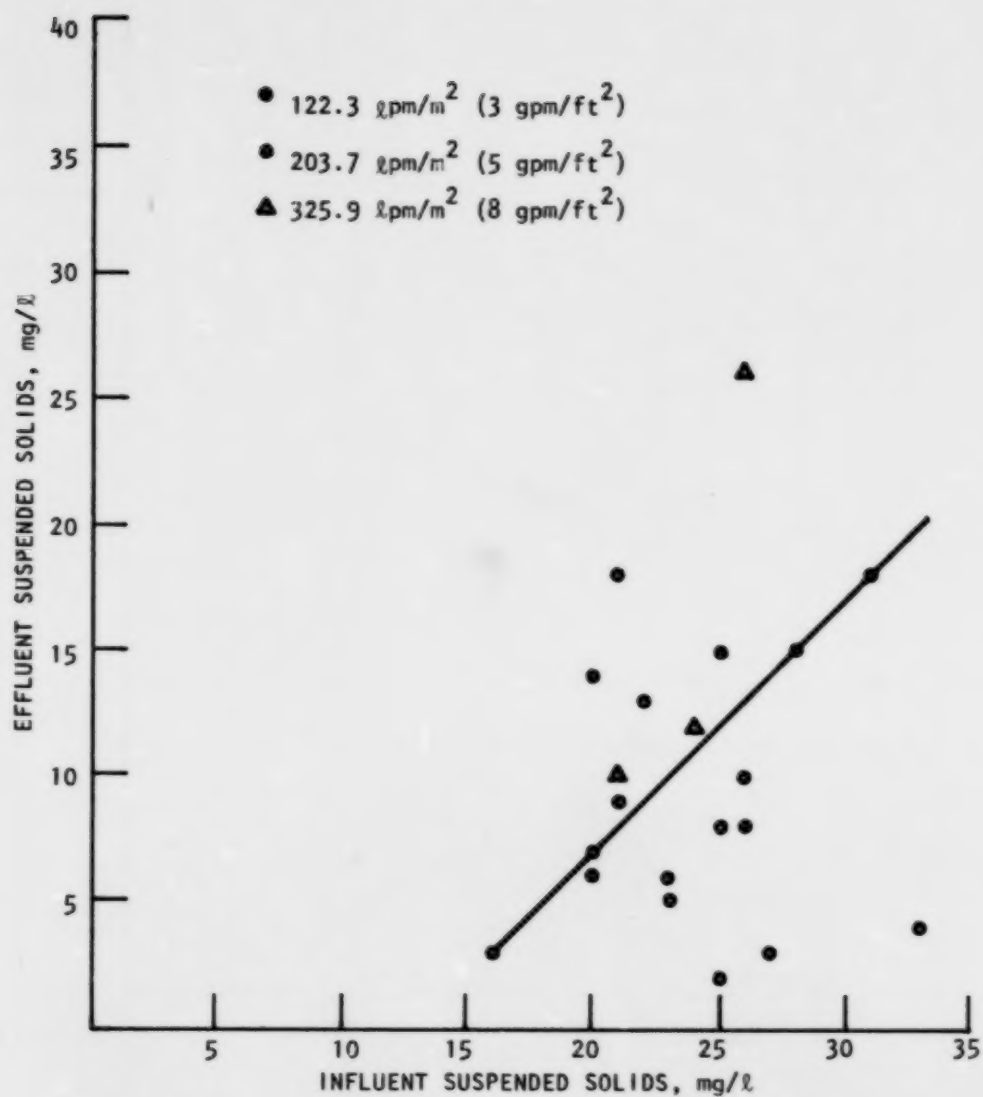


Figure 42. Dual media filter relationship of effluent to influent suspended solids for chemically treated blast furnace blowdown.

Table 31 summarizes the average solids loading for various filtration rates. These data are presented graphically in Figure 43 to show the relationship between solids loading and filtration rate.

TABLE 31. SUMMARY OF DUAL MEDIA FILTER DATA FOR FILTRATION WITH CHEMICALLY PRETREATED BLAST FURNACE BLOWDOWN

Filtration rate		Average solids loading	
ℓ/min/m ²	gpm/ft ²	kg SS/m ² /m H.L.	lb SS/ft ² /ft H.L.
122.2	3	1.04	0.065
203.7	5	0.55	0.036
325.9	8	0.35	0.022

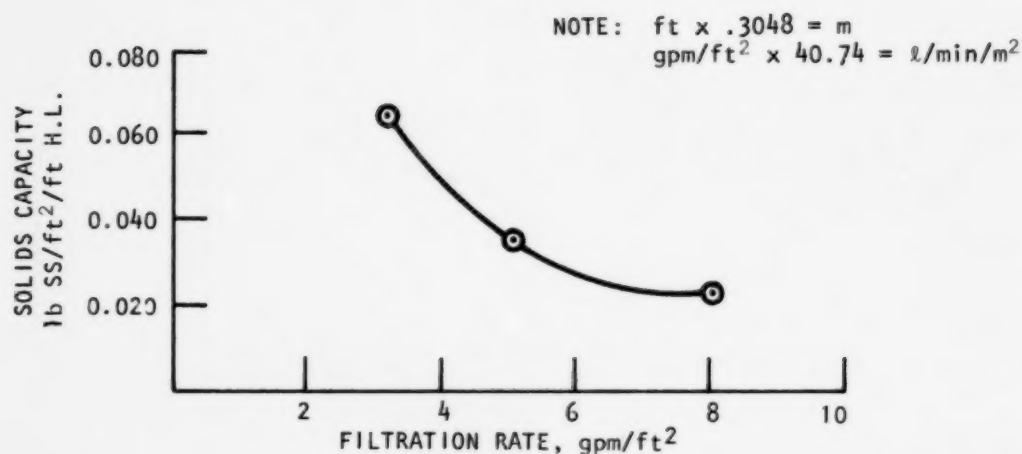


Figure 43. Dual media filter solids capacity with chemically pretreated blast furnace blowdown.

The data presented in the table and graph indicated that solids loading at different flow rates are quite variable, ranging from a high of 1.04 kg SS/m²/m H.L. (0.065 lb SS/ft²/ft H.L.) at 122.2 ℓ/min/m² (3 gpm) to a low of 0.35 kg SS/m²/m H.L. (0.022 lb SS/ft²/ft H.L.) at 325.9 ℓ/min/m² (8 gpm).

Basis for optimum filtration rate selection--Figure 44 shows that the highest effluent quality was obtained at a filtration rate of 203 l/min/m^2 (5 gpm/ft^2). Assuming an average feed suspended solids of 27 mg/l and a required effluent quality of 12 mg/l suspended solids, Figure 44 shows that a filtration rate between 162 l/min/m^2 (4 gpm/ft^2) and 244 l/min/m^2 (6 gpm/ft^2) may be utilized. Figure 45 shows that operation at a filtration rate of 203 l/min/m^2 (5 gpm/ft^2) with 9.1 m (30 ft) of headloss would result in a run time of 29 hours. For the same rate but with a terminal headloss of 10 ft , Figure 45 reveals a run time of 9.5 hours. This run time is below the normally accepted minimum run time, therefore the filtration rate should be decreased to permit a longer run time.

NOTE: $\text{gpm/ft}^2 \times 40.74 = \text{l/min/m}^2$

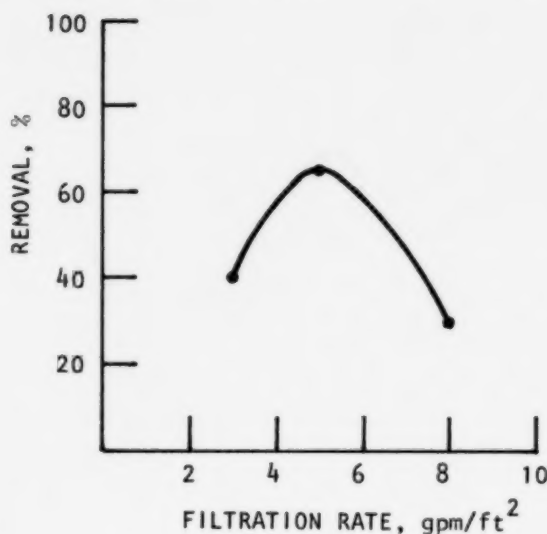


Figure 44. Effect of filtration rate on solids removal of chemically pretreated blast furnace blowdown.

For design purposes, based on effluent requirements, backwash frequency and equipment design, a filtration rate of 203 l/min/m^2 (5 gpm/ft^2) was chosen for the pressure filter and a filtration rate of 183 l/min/m^2 (4.5 gpm/ft^2) was chosen for the gravity filter.

Headloss Profile

During each filter run, headloss profiles were taken to indicate how solids

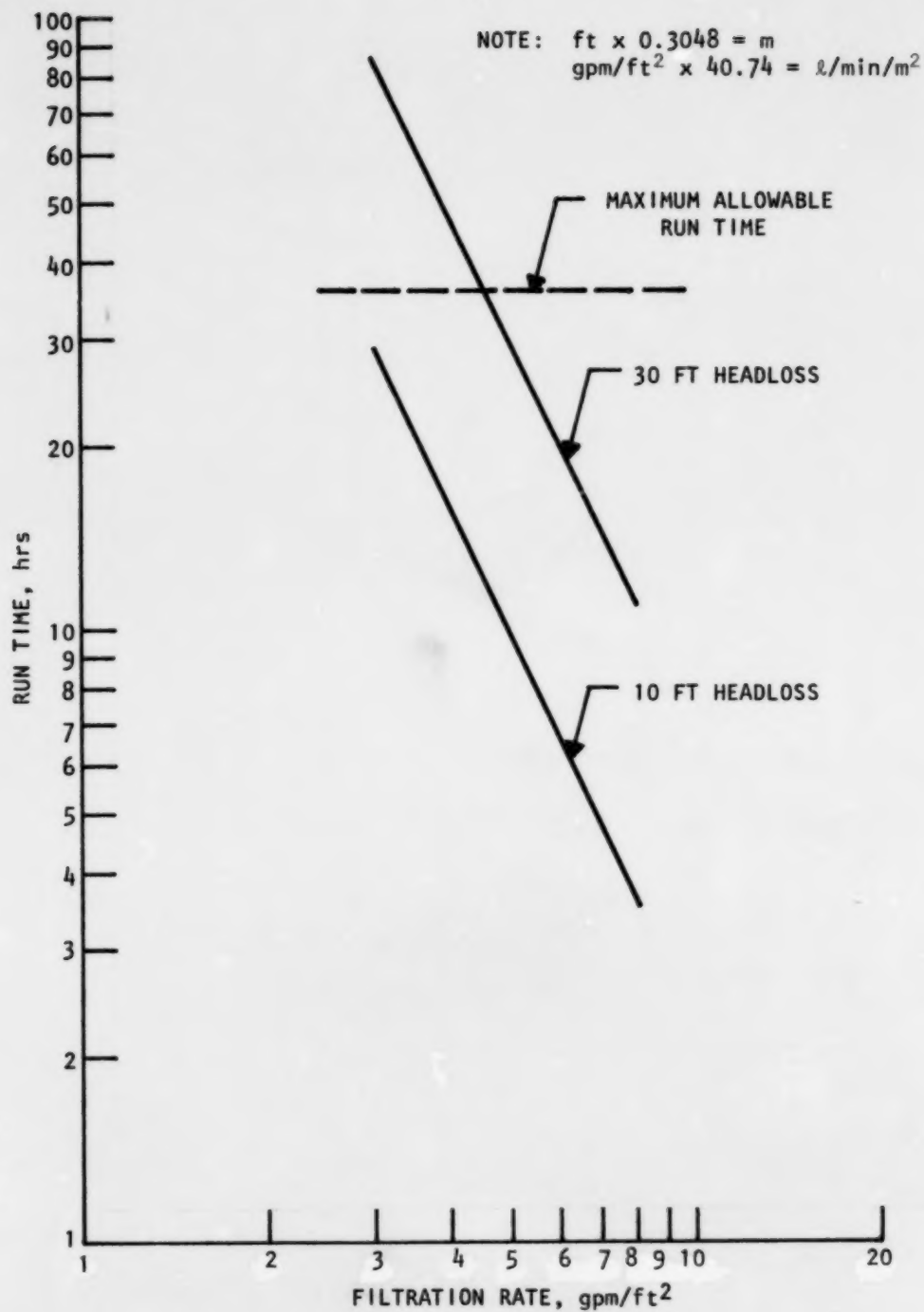
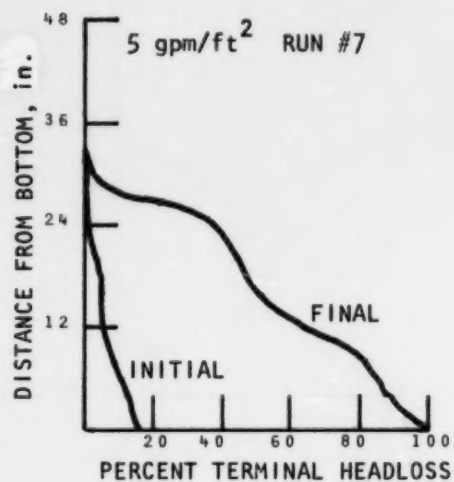
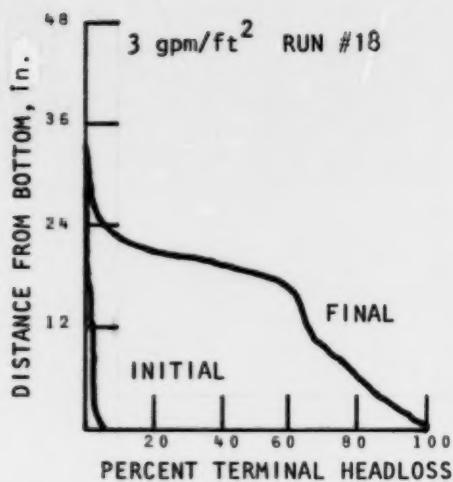


Figure 45. Dual media filter run time -
chemically pretreated blowdown.



NOTE: in. $\times$ 2.54 = cm
 gpm/ft² $\times$ 40.74 = l/min/m²

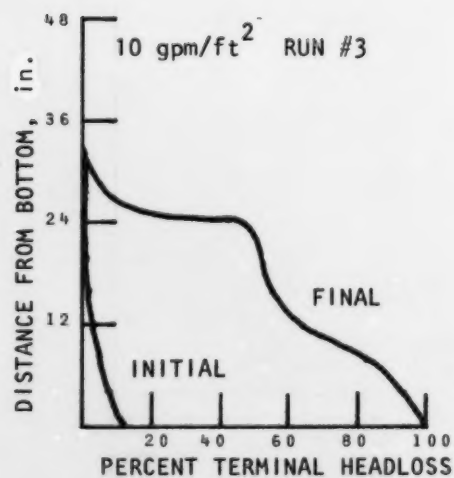
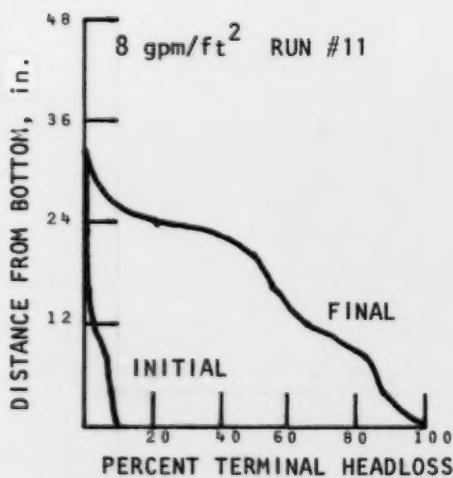
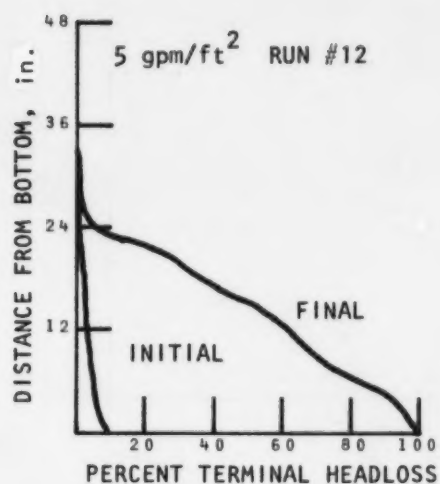
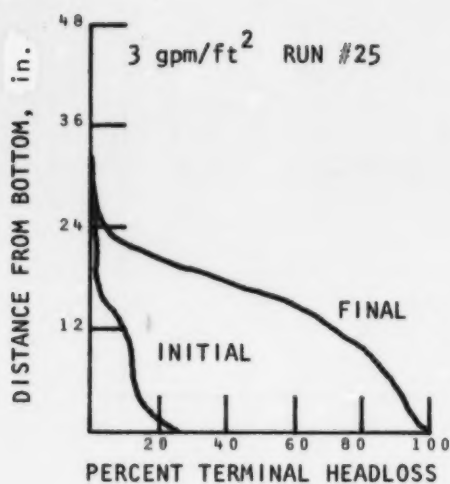


Figure 46. Headloss profile with raw blast furnace blowdown.



NOTE: in. $\times$ 2.54 = cm
 gpm/ft² $\times$ 40.74 = l/min/m²

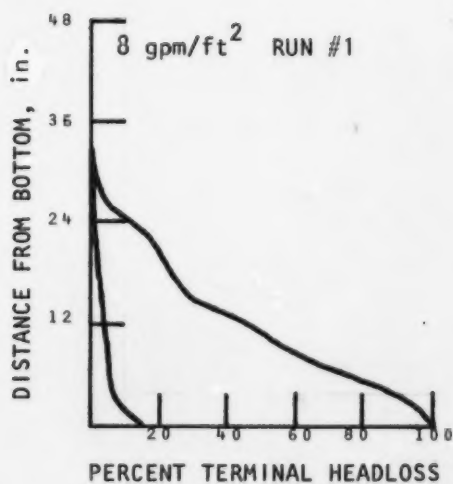


Figure 47. Headloss profile with chemically pretreated blowdown.

were deposited in the filter media. Typical headloss profiles are shown in Figures 46 and 47 for the filtration rates tested.

The flatness of the profile curve for the coal media indicates the occurrence of surface filtration. A modest slope of the profile line indicates the presence of in-depth filtration. A clearly defined sand/anthracite interface was not observed.

Backwash

Throughout the pilot program the same backwash procedure was used. There was no observable deterioration of filter effectiveness as measured by the initial headloss or other performance parameters. A typical washout curve of suspended solids shown in Figure 48 indicates adequate removal of solids when the wash rate was $530 \text{ l/m}^2/\text{backwash}$ ($130 \text{ gal./ft}^2/\text{backwash}$).

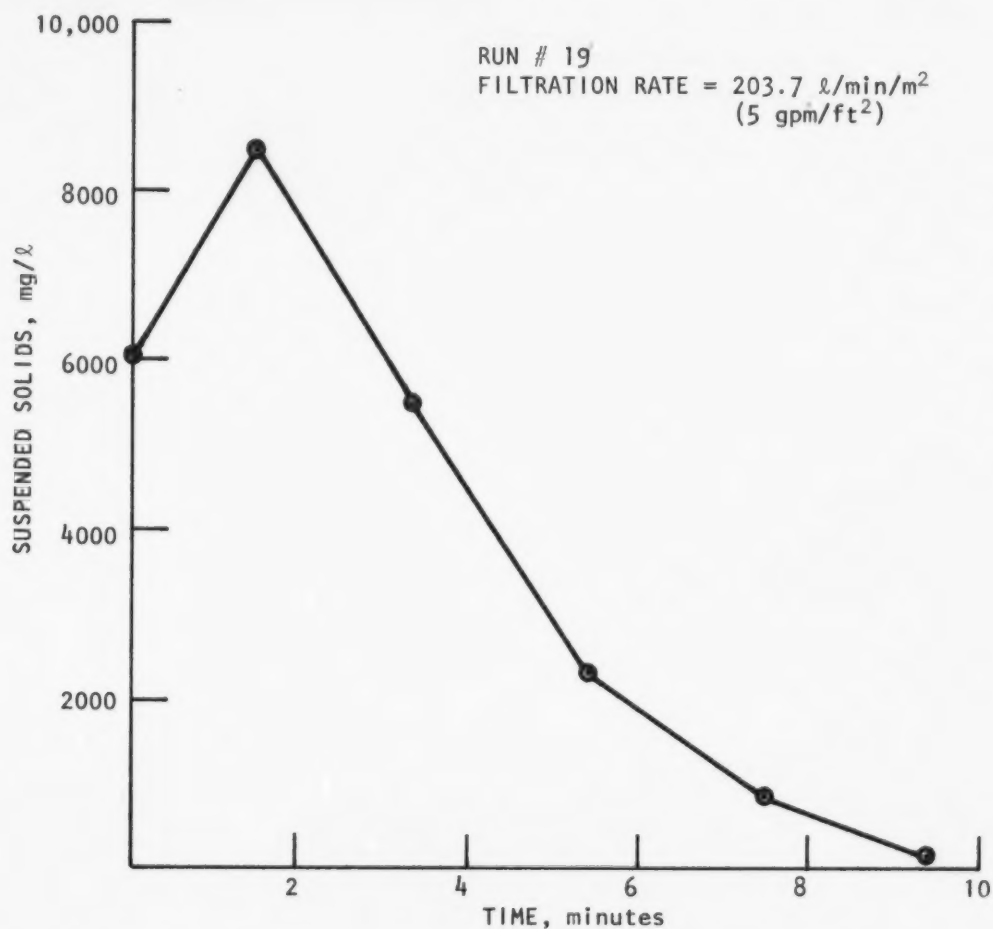


Figure 48 . Typical suspended solids washout curve.

Cost and Space Requirements

Design Parameters--

From the preceding discussions, design parameters were selected for filtration of secondary effluent and alum clarified effluent. Tables 32, 33, 34 and 35 summarize the design basis used for cost estimates. The gravity filter mode is operated at 10 feet headloss while pressure filtration is operated at 30 feet headloss.

Gravity filtration will be based on "variable declining rate" filter operating mode. Pressure filter operation will be based on a constant-head rate.

Pressure filters will be carbon steel vertical type 100 lb code vessels with false bottoms, stainless steel diffusers, carbon steel piping, automatic butterfly valves, air blower and required filter feed pumps. (See Figure 49).

Gravity filters will be carbon steel tanks with automatically operated butterfly valves, underdrains of the head-lateral type, and an air blower. In addition two backwash pumps and a holding tank will be provided for the backwash system. (See Figure 50).

Both systems will be capable of being backwashed first with air at $1.5 \text{ m}^3/\text{min}/\text{m}^2$ (5 scfm/ft²) and then with water at $610 \text{ l}/\text{min}/\text{m}^2$ (15.0 gpm/ft²).

Cost Considerations--

Having selected appropriate design parameters, dual media filtration costs were computed for flows of $946 \text{ m}^3/\text{day}$ (0.25 mgd), $2,838 \text{ m}^3/\text{day}$ (0.75 mgd), and $5,677 \text{ m}^3/\text{day}$ (1.5 mgd). Cost comparisons included gravity versus pressure filtration for raw blast furnace blowdown and chemically pretreated blast furnace blowdown.

Equipment costs--Costs for filter components were based on vendor quotations, and from a recent edition of the Dodge Manual for Building Construction, Pricing and Scheduling (6). Equipment costs include:

1. Filter vessels and components.
2. Backwash reservoir.
3. Valves, actuators and controls for automatic operation.
4. Media.
5. Pumps and motor control center.
6. Air compressor for air scour.
7. Control console.
8. Interconnecting piping between filter vessels.
9. A proportion of the clarifier costs was allowed for treatment of the backwash.

TABLE 32. DUAL MEDIA PRESSURE FILTER DESIGN DATA RAW BLAST FURNACE BLOWDOWN

Type of waste	Raw blast furnace blowdown	
Design flow, m ³ /day (mgd)	946 (0.25)	2838 (0.75)
Filter type	pressure	pressure
Design influent, mg/l S.S.	84	84
Design effluent, mg/l S.S.	15	15
Filtration rate, l/min/m ² (gpm/ft ²)	203.7 (5)	203.7 (5)
Area required, m ² (ft ²)	3.3 (35.7)	9.9 (107)
Filter diameter, m (ft)	1.22 (4.0)	2.06 (6.75)
Filter height, m (ft)	2.13 (7.0)	2.44 (8.0)
Filter configuration	vertical	vertical
Number of filters	3	3
Area per filter, m ² (ft ²)	1.2 (12.6)	3.3 (35.8)
Backwash usage, %	1.4	1.4
Backwash cycle duration, min.	30	30
Filter cycle time, hr.	36.5	36.5
Depth of silica sand, cm (in.)	50.8 (20)	50.8 (20)
Depth of anthracite, cm (in.)	45.7 (18)	45.7 (18)
Depth of gravel, cm (in.)	10.2 (4)	10.2 (4)

TABLE 33. DUAL MEDIA GRAVITY FILTER DESIGN DATA RAW BLAST FURNACE BLOWDOWN

Type of Waste	Raw Blast Furnace Blowdown		
Design flow, m ³ /day (mgd)	946 (0.25)	2838 (0.75)	5677 (1.5)
Filter type	Gravity	Gravity	Gravity
Design influent, mg/l S.S.	84	84	84
Design effluent, mg/l S.S.	15	15	15
Filtration rate, l/min/m ² (gpm/ft ²)	203.7 (5)	203.7 (5)	203.7 (5)
Area required, m ² (ft ²)	3.3 (36)	10.0 (108)	20.0 (215)
Filter diameter, m (ft)	1.22 (4.0)	2.06 (6.75)	2.90 (9.5)
Filter height, m (ft)	3.5 (11.5)	3.5 (11.5)	3.5 (11.5)
Filter configuration	Vertical	Vertical	Vertical
Number of filters	3	3	3
Area per filter, m ² (ft ²)	1.2 (12.6)	3.3 (35.8)	6.6 (70.8)
Backwash usage, %	1.7	1.7	1.7
Backwash cycle duration, min.	30	30	30
Filter cycle time, hr.	30.5	30.5	30.5
Depth of silica sand, cm (in.)	50.8 (20)	50.8 (20)	50.8 (20)
Depth of anthracite, cm (in.)	40.6 (16)	40.6 (16)	40.6 (16)
Depth of gravel, cm (in.)	48.3 (19)	48.3 (19)	48.3 (19)

TABLE 34. DUAL MEDIA PRESSURE FILTER DESIGN DATA
CHEMICALLY PRETREATED BLAST FURNACE BLOWDOWN

Type of Waste	Chemically Pretreated Blast Furnace Blowdown		
	946 (0.25)	2838 (0.75)	5677 (1.5)
Design flow, m ³ /day (mgd)	Pressure	Pressure	Pressure
Filter type	27	27	27
Design influent, mg/ℓ S.S.	12	12	12
Design effluent, mg/ℓ S.S.	203.7 (5)	203.7 (5)	203.7 (5)
Filtration rate, ℓ/min/m ² (gpm/ft ²)	3.3 (36)	10.0 (108)	20.0 (215)
Area required, m ² (ft ²)	1.22 (4.0)	2.06 (6.75)	2.90 (9.5)
Filter diameter, m (ft)	2.13 (7.0)	2.44 (8.0)	2.74 (9.0)
Filter height, m (ft)	Vertical	Vertical	Vertical
Filter configuration	3	3	3
Number of filters	1.2 (12.6)	3.3 (35.8)	6.6 (70.8)
Area per filter, m ² (ft ²)	1.7	1.7	1.7
Backwash usage, %	30	30	30
Backwash cycle duration, min.	29.5	29.5	29.5
Filter cycle time, hr.	50.8 (20)	50.8 (20)	50.8 (20)
Depth of silica sand, cm (in.)	45.7 (18)	45.7 (18)	45.7 (18)
Depth of anthracite, cm (in.)	10.2 (4)	10.2 (4)	10.2 (4)
Depth of gravel, cm (in.)			

TABLE 35. DUAL MEDIA GRAVITY FILTER DESIGN DATA
CHEMICALLY PRETREATED BLAST FURNACE BLOWDOWN

Type of Waste	Chemically Pretreated Blast Furnace Blowdown		
Design flow, m ³ /day (mgd)	946 (0.25)	2838 (0.75)	5677 (1.5)
Filter type	Gravity	Gravity	Gravity
Design influent, mg/ℓ S.S.	27	27	27
Design effluent, mg/ℓ S.S.	12	12	12
Filtration rate, ℓ/min/m ² (gpm/ft ²)	183.3 (4.5)	183.3 (4.5)	183.3 (4.5)
Area required, m ² (ft ²)	3.9 (42)	11.6 (125)	23.2 (250)
Filter diameter, m (ft)	1.30 (4.25)	2.21 (7.25)	3.12 (10.25)
Filter height, m (ft)	2.13 (7.0)	2.44 (8.0)	2.74 (9.0)
Filter configuration	Vertical	Vertical	Vertical
Number of filters	3	3	3
Area per filter, m ² (ft ²)	1.3 (14.2)	3.8 (41.3)	7.7 (82.5)
Backwash usage, %	3.8	3.8	3.8
Backwash cycle duration, min.	30	30	30
Filter cycle time, hr.	13.2	13.2	13.2
Depth of silica sand, cm (in.)	50.8 (20)	50.8 (20)	50.8 (20)
Depth of anthracite, cm (in.)	40.6 (16)	40.6 (16)	40.6 (16)
Depth of gravel, cm (in.)	48.3 (19)	48.3 (19)	48.3 (19)

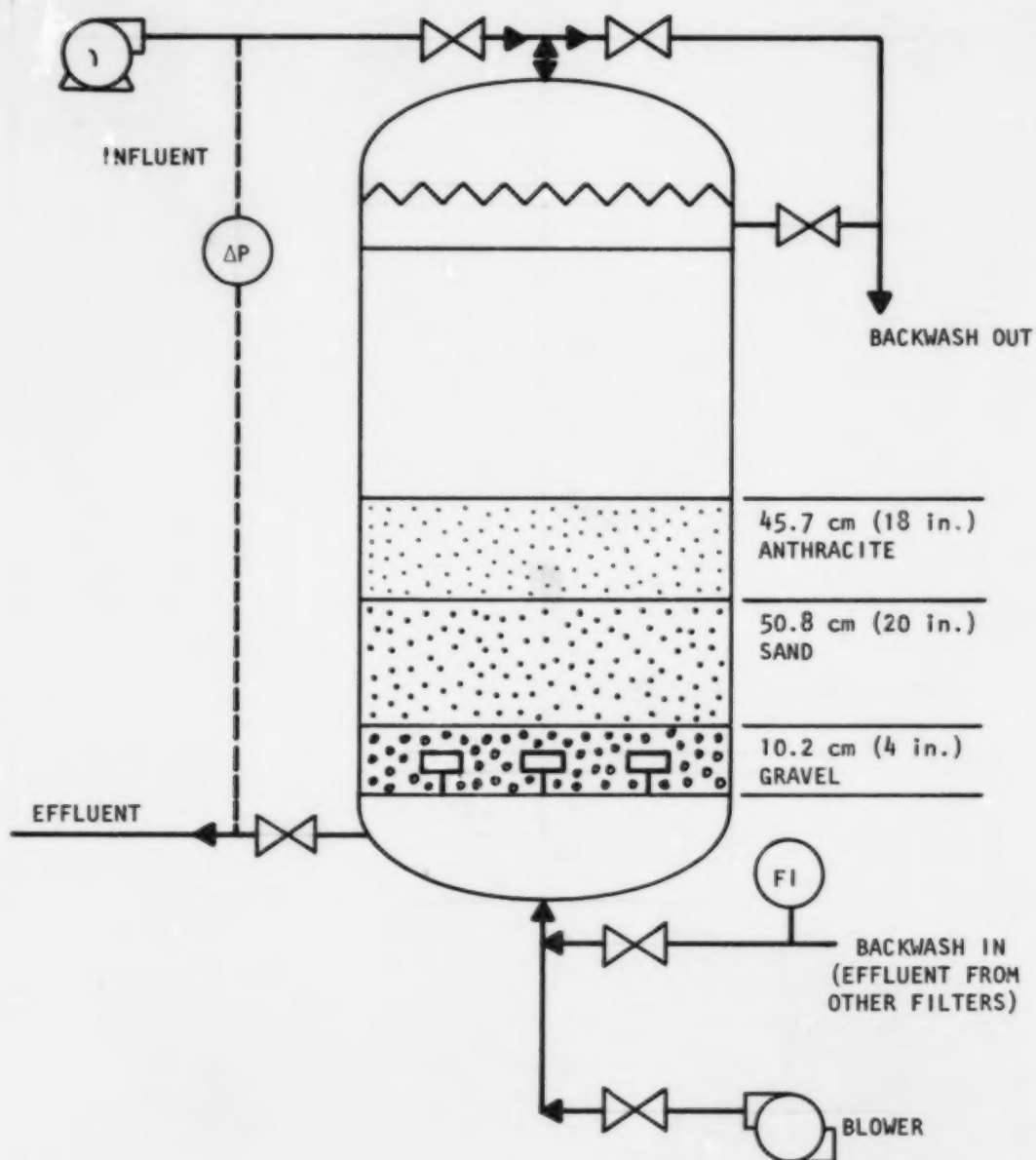


Figure 49. Front view of pressure filter.

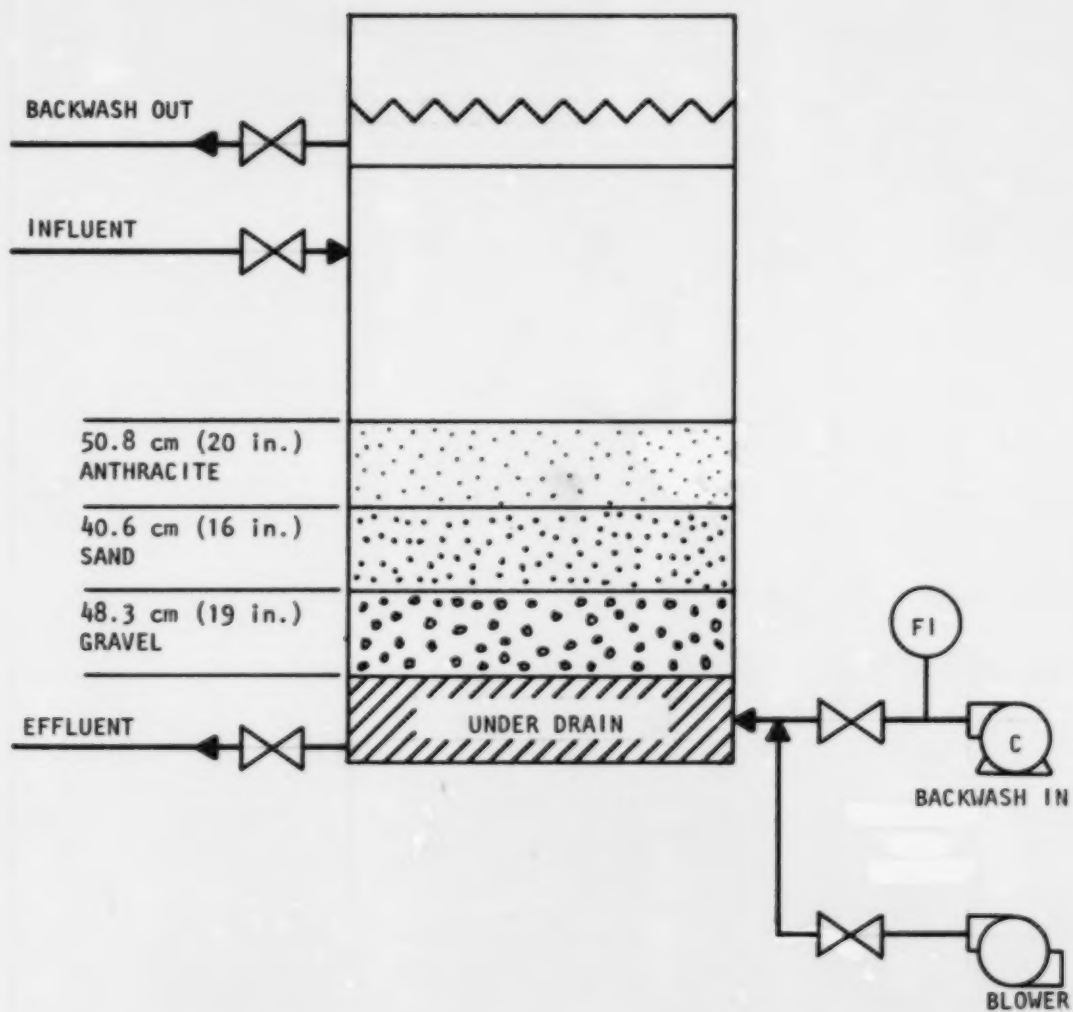


Figure 50. Front view of gravity filter.

Discussion of Dual Media Filtration Costs--

Tables 36 and 37 summarize the cost data for dual media filtration. The costs are broken out for major equipment items and equipment subtotal as well as major operating cost items and yearly operating subtotal. These costs were then tabulated to show yearly operating cost and cost per 3,785 liters (one thousand gallons) design flow.

Operating costs--Components of the yearly operating cost include power, operation labor, maintenance labor, maintenance material and amortization of capital. Items included in the analysis are discussed as follows:

Electrical: For all filters, the costs of operating the backwash pumps and air scour compressors were based on the backwash frequency and the cycle duration. Electrical costs for pressure filtration include power costs for the feed pumps operating at the design flow, 24 hours per day. Power costs were assumed to be 3.0¢/KWH.

Operation Labor: Operation of the 946 m³/day (0.25 mgd), 2,838 m³/day (0.75 mgd), and the 5,677 m³/day (1.5 mgd) plans was expected to require 40, 40 and 50 percent operator attention respectively. The first shift operator was assumed to be skilled labor at \$18,000 per year, while the second and third shift operators would be unskilled at \$16,000 per year.

Maintenance Material: Included are allowances for replacement parts and replacement of lost media.

Capital Amortization: An equipment service life of twenty years, amortized at 6.875 percent per year was assumed.

Items Not Included: Examples of items not included in this analysis are; fringe benefits, supervision, overhead, laboratory support, insurance and administration expenses. It is reasonable to expect such factors to double the total operating costs.

Tables 36 and 37 show the most sensitive component to the various process changes is the power cost. The added pumps required for pressure filtration increase the power costs by a factor of ten over gravity filtration. To a lesser extent, the frequency of backwash is reflected in power costs.

Labor costs are somewhat independent of filter size and flow loading and can be considered fixed operating costs. Material for maintenance is a function of total capital costs.

The yearly operating total shows pressure filtration costs more to operate than gravity filters. Overall cost effectiveness comparisons are calculated on the basis of equivalent annual values which include both capital amortization and operating costs. Results shown in Tables 36 and 37 are expressed as total yearly costs and costs per 3,785 liters (one thousand gallons) design flow. The comparison shows that gravity filtration for both raw and chemically pretreated blast furnace blowdown is more cost-effective with increasing design flow.

TABLE 36. DUAL MEDIA FILTER COST SUMMARY
RAW BLAST FURNACE BLOWDOWN^a

Filter Type	946 (0.25)		2,838 (0.75)		5,677 (1.5)	
	Gravity	Pressure	Gravity	Pressure	Gravity	Pressure
Design flow, m ³ /day (mgd)						
Capital Cost ^b						
Filtered components	\$ 19,000	\$ 28,000	\$ 35,000	\$ 52,000	\$ 53,000	\$ 78,000
Feed pumps	-	4,000	-	6,000	-	9,000
Backwash pump & tank	6,000	-	9,000	-	15,000	-
Upstream treatment of backwash	2,000	1,000	2,000	2,000	3,000	2,000
Delivered Equipment Total	\$ 27,000	\$ 33,000	\$ 46,000	\$ 60,000	\$ 71,000	\$ 89,000
Equipment installation	11,000	13,000	18,000	24,000	29,000	36,000
Installed Equipment Total	\$ 38,000	\$ 46,000	\$ 64,000	\$ 84,000	\$ 100,000	\$ 125,000
Process piping	12,000	14,000	19,000	25,000	30,000	38,000
Instrumentation	2,000	2,000	3,000	4,000	5,000	6,000
Capital Equipment Total	\$ 52,000	\$ 62,000	\$ 86,000	\$ 113,000	\$ 135,000	\$ 169,000
Facilities & Fees	111,000	133,000	185,000	242,000	291,000	364,000
Total Plant Cost	\$163,000	\$195,000	\$271,000	\$355,000	\$426,000	\$533,000
Operating Cost						
Electrical	\$ -	\$ 2,000	\$ -	\$ 5,000	\$ 1,000	\$ 9,000
Operators	20,000	20,000	20,000	20,000	25,000	25,000
Maintenance Labor & Materials	4,000	5,000	7,000	8,000	10,000	11,000
Yearly Operating Total	\$ 24,000	\$ 27,000	\$ 27,000	\$ 33,000	\$ 36,000	\$ 45,000
Capital Amortization (20 yr life; 6.875%)	15,000	18,000	25,000	33,000	40,000	50,000
Total Yearly Cost	\$ 39,000	\$ 45,000	\$ 52,000	\$ 66,000	\$ 76,000	\$ 95,000
c/3,785 gal. (c/1,000 gal.)	42.7	49.3	19.0	24.1	13.9	17.4
Land Requirement m ² (acres)	100 (0.25)	70 (.017)	160 (.040)	110 (0.27)	250 (.062)	180 (.044)

a. If uniform flow to the filter systems can not be maintained, an equalization basin may be required.
There has been no cost allowance for equalization in the above designs.
b. See Appendix Tables A-4 and A-5 for basis of estimate.

TABLE 37. DUAL MEDIA FILTER COST SUMMARY
CHEMICALLY PRETREATED BLAST FURNACE BLOWDOWN

Design flow, m ³ /day (mgd)	946 (0.25)		2,838 (0.75)		5,677 (1.5)	
Filter Type	Gravity	Pressure	Gravity	Pressure	Gravity	Pressure
Capital Cost^a						
Filtered components	\$ 19,000	\$ 28,000	\$ 35,000	\$ 52,000	\$ 53,000	\$ 78,000
Feed pumps	-	4,000	-	6,000	-	9,000
Backwash pump & tank	6,000	-	9,000	-	15,000	-
Upstream treatment of backwash	2,000	1,000	2,000	2,000	3,000	2,000
Delivered Equipment Total	\$ 27,000	\$ 33,000	\$ 46,000	\$ 60,000	\$ 71,000	\$ 89,000
Equipment installation	11,000	13,000	18,000	24,000	29,000	36,000
Installed Equipment Total	\$ 38,000	\$ 46,000	\$ 64,000	\$ 84,000	\$ 100,000	\$ 125,000
Process piping	12,000	14,000	19,000	25,000	30,000	38,000
Instrumentation	2,000	2,000	3,000	4,000	5,000	6,000
Capital Equipment Total	\$ 52,000	\$ 62,000	\$ 86,000	\$ 113,000	\$ 135,000	\$ 169,000
Facilities & Fees	111,000	133,000	185,000	242,000	291,000	364,000
Total Plant Cost	\$163,000	\$195,000	\$271,000	\$355,000	\$426,000	\$533,000
Operating Cost						
Electrical	\$ -	\$ 2,000	\$ -	\$ 5,000	\$ 1,000	\$ 9,000
Operators	20,000	20,000	20,000	20,000	25,000	25,000
Maintenance Labor & Materials	4,000	5,000	7,000	8,000	10,000	11,000
Yearly Operating Total	\$ 24,000	\$ 27,000	\$ 27,000	\$ 33,000	\$ 36,000	\$ 45,000
Capital Amortization (20 yr life; 6.875%)	15,000	18,000	25,000	33,000	40,000	50,000
Total Yearly Cost	\$ 39,000	\$ 45,000	\$ 52,000	\$ 66,000	\$ 76,000	\$ 95,000
¢/3,785L (¢/1,000 gal.)	42.7	49.3	19.0	24.1	13.9	17.4
Land Requirement m ² (acres)	100 (0.25)	70 (.017)	160 (.040)	110 (0.27)	250 (.062)	180 (.044)

a. See Appendix Tables A-4 and A-5 for basis of estimate.
If uniform flow to the filter systems can not be maintained, an equalization basin may be required.
There has been no cost allowance for equalization in the above designs.

Equipment costs--Process considerations determine the filter area required and the frequency of backwash, so equipment costs reflect these differences. Comparing the cost of filter components it is apparent that higher costs are associated with pressure filtration than gravity filtration. The reasons for this difference are the higher cost for the pressure vessels and the added expense for feed pumps. The cost for these items was greater than the increased cost incurred for the backwash system of the gravity filters. The pressure filters employ their feed pumps for backwashing and therefore no additional costs are incurred.

Gravity filtration costs for chemically pretreated blast furnace blowdown were slightly higher than for raw blast furnace blowdown. This was because a lower filtration rate (or larger filters) had to be employed to obtain satisfactory run lengths. This also resulted in increased size of upstream clarification equipment needed to treat the backwash water.

Absolute cost considerations--The cost estimates presented in Tables 36 and 37 are internally consistent and suitable for comparison purposes within the scope of this report. It should be realized that the costs are limited to items listed on page 121 and do not represent the absolute value of total installed cost. The intent is to compare the various alternatives with detail not available in published cost curves. For this reason it is recognized that the absolute costs reported are less than the published costs of granular media filtration.

Land requirements--Also tabulated in Tables 36 and 37 are the space requirements for the various options. For the pressure filter system the filters were located along one wall with adequate spacing between filters and walls to permit easy access to interconnecting piping and valving. The remainder of the building included a central aisle way and the opposite wall included space for the feed pumps, blower and control panel. For the gravity filter systems the building had to be enlarged to include the required holding tank.

RESULTS OF THE MAGNETIC FILTER STUDY

Introduction

Magnetic filtration tests were performed on the blast furnace blowdown from December 16, 1978 through March 27, 1979. During December and January, filtration of blowdown chemically pretreated with alum and polymer were conducted.

Tables E-1 and E-3 in the Appendix present the suspended solids analysis of all 24 hour composites taken during the study. Included in these tables are information on influent and effluent quality and solids loading for each run of the magnetic filter. Two identifiable types of feed were filtered as shown in the summary Tables 38 and 39. These included raw and chemically pretreated blowdown.

Filtration of Raw Blast Furnace Blowdown

Magnetic filtration results of the raw blast furnace blowdown are summarized in Table 38. Filtration removed moderate quantities of suspended solids (70 percent removal) and turbidity (32 percent removal).

TABLE 38. MAGNETIC FILTRATION OF RAW BLAST FURNACE BLOWDOWN-INFLUENT VS EFFLUENT QUALITY

	pH	SS, mg/l	Temp. °C	Turbidity, FTU
			Influent	
Average	-	81	22	40
Maximum	7.8	97	24	59
Minimum	6.5	51	21	31
			Effluent	
Average	-	24	22	27
Maximum	7.8	32	24	40
Minimum	6.5	8	21	9

Filtration of Chemically Pretreated Blast Furnace Blowdown

Table 39 shows the average, maximum and minimum influent and effluent characteristics when chemically pretreated blowdown was filtered. Filtration removed only small quantities of suspended solids (32 percent removal) and turbidity (37 percent removal).

TABLE 39. MAGNETIC FILTRATION OF CHEMICALLY PRETREATED BLOWDOWN-INFLUENT VS EFFLUENT QUALITY

	pH	SS, mg/l	Temp., °C	Turbidity, FTU
			Influent	
Average	-	28	23	10.3
Maximum	7.2	31	23	17
Minimum	6.8	25	22	5.3
			Effluent	
Average	-	19	23	6.5
Maximum	7.2	24	23	11
Minimum	6.8	11	22	3.3

Filtration of Chemically Pretreated (with Magnetite) Blast Furnace Blowdown

Table 40 shows the influent and effluent characteristics when chemically pretreated (with magnetite) blowdown was filtered. During the test, 200 mg/l alum, 0.5 mg/l polymer and 200 mg/l magnetite were added to the wastewater in an attempt to improve effluent quality. Overall removal of suspended solids (4 percent removal) and turbidity (28 percent removal) was lower than expected. Chemical dosages were based on laboratory screening tests.

TABLE 40. MAGNETIC FILTRATION OF CHEMICALLY PRETREATED (WITH MAGNETITE) BLOWDOWN-INFLUENT VS EFFLUENT QUALITY

	pH	SS, mg/l	Temp., °C	Turbidity, FTU
Influent	6.9	608	-	25
Effluent	6.9	581	-	18

Data Collection

In addition to overall performance data, the effect of specific operating parameters was determined for each filter run. Since suspended solids removal is the most important measure of filter performance, this parameter was monitored during each run in addition to the filtration rate, headloss and run time.

Accumulation of solids causes increasing pressure drop across the matrix. During any pilot study, the feed suspended solids varies from run to run and rarely equals the design conditions. Therefore, the most meaningful parameter of filter performance is the amount of solids trapped by the filter per unit area per unit headloss. This is usually reported in terms of kilograms suspended solids/m²/m headloss (pounds suspended solids/sq ft/ft H.L.).

Appendix Tables E-1, E-2, E-3 and E-4 list all filter runs where complete information on filtration rate, headloss, run time, suspended solids, turbidity, etc. were available. Also tabulated are the calculated solids loading capacities of each run. The tables separate runs with raw blast furnace blowdown (Tables E-1, E-2) from those with chemically pretreated blowdown Tables (E-3, E-4).

Discussion of Results

All test runs behaved in a similar manner with the best quality effluent being obtained near the beginning of the run. Usually at the very beginning of the run, the initial effluent was of lower quality until some of the material dislodged during backwashing was rinsed out. From that point,

there was a gradual deterioration of effluent turbidity quality. Operating the unit at the manufacturer's recommended flow rate of $2,648 \text{ l/min/m}^2$ (65 gpm/ft^2) resulted in about 50 percent turbidity removal initially with a pressure drop of 0.21 kg/cm^2 (3 psi). With time, the pressure drop gradually increased and the effluent turbidity removal decreased to about 25 percent after 1-2 hours. At this point the pressure drop and turbidity level reached a plateau and only gradually increased with time. However, the effluent quality was only marginally acceptable at that time, so nearly all runs were less than 4 hours.

Since only composite samples were taken for suspended solids, it is not possible to plot the percent suspended solids removal with time, and it can only be inferred from what was observed from the turbidity grab samples.

A typical plot of turbidity versus time is shown in Figure 51 which illustrates the previous discussion.

For the raw blast furnace blowdown, the suspended solids removal improved with decreasing flow. A similar trend was observed for the chemically pretreated blowdown, but additional sample points would be required to verify the trend.

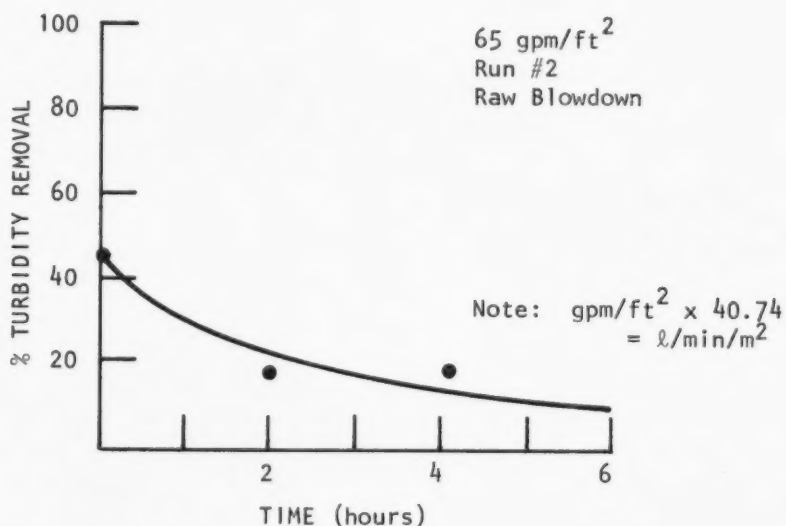


Figure 51. Turbidity removal versus time for magnetic filtration.

The one run conducted using magnetite seed with the chemically pretreated blowdown gave disappointing results. The filter did not prove effective in removing the magnetite floc from the wastewater with the net result being that this effluent was of poorer quality than "non-magnetite" effluent. Although encouraging results have been achieved by other researchers (12,13) test results using an optimized magnetite:alum:polymer ratio were not promising for the particular blast furnace wastewater investigated. Additional testing on a different blast furnace water is certainly warranted.

Backwash

Throughout the pilot program the same backwash procedure was used. There was no observable deterioration of filter effectiveness as measured by initial headloss on other performance parameters. Midway through the program the filter's media (steel wool) was changed, with no change in performance observed. The filter was backwashed at 18.9 ℓ/min . (5 gpm) for 90 seconds throughout the study. An analysis of a typical washout curve in Figure 52 indicates adequate removal of solids after about 30 seconds.

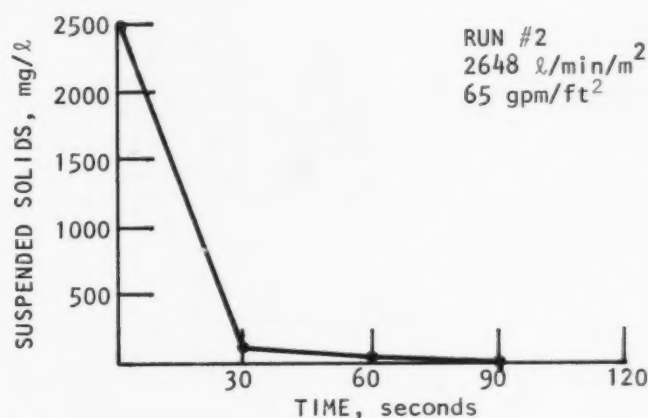


Figure 52. Typical suspended solids washout from magnetic filter during backwash.

Equipment Costs

For raw blast furnace blowdown the magnetic filter removed 70 percent of the suspended solids and 32 percent of the turbidity while the dual media filter removed 83 percent and 75 percent, respectively. For the chemically pretreated wastewater, the magnetic filter removed 32 percent of the suspended solids and 37 percent of the turbidity while the dual media filter removed 44 percent and 66 percent respectively.

The dual media and magnetic filter were investigated to provide pretreatment to downstream processes. The principal parameters of concern were suspended

solids and turbidity. Therefore, based on the above discussion, dual media filtration consistently provided a better quality effluent and is recommended for a full-scale system at this plant. Due to the distinct advantage seen in dual-media filtration, a cost analysis will not be undertaken for magnetic filtration.

RESULTS OF THE REVERSE OSMOSIS STUDY

Membranes

The reverse osmosis (RO) testing program was performed on the blast furnace wastewater from February 28 to March 26, 1979. During Phase 1 of the testing, the system was operated at 31.9 - 33.6 percent product water recoveries. The system operated at 71.9 - 81.0 percent water recoveries during Phase 2 of the program.

TFC Model 4600 PA membranes, purchased from Fluid Systems Division, Universal Oil Products Inc. (UOP), were used in the program. The characteristics of this membrane are presented in Table 41. These membranes were selected to evaluate the new technology represented by the polyamide (PA) membranes. The PA membranes have several advantages over cellulose acetate membranes (14), including:

1. Higher permissible feedwater temperature.
2. Broader operating pH range [although a pH range of 4.0 - 6.0 is recommended, PA membranes have proven stable in a range of 3.0 - 12.0 (14)].
3. Higher inorganic and organic rejections.
4. Greater flux per membrane.
5. Exhibits a resistance to compaction at high temperature and pressure.

The RO unit consisted of one pressure vessel containing three membranes in series.

Pretreatment

The feed to the RO system was chemically treated, clarified, filtered, and pH adjusted blowdown from the blast furnace scrubber system. Acid addition, using concentrated hydrochloric acid, was required to prevent CaCO_3 scaling in the membranes. An automatic pH controller and acid pump were utilized to maintain the desired range. The normal pH range of 7.10 - 8.30 was lowered to 6.0 - 6.2. The desired pH range was determined using the concentrations of the various anions and cations, expected product water recoveries, and the Langlier Index (15).

A summary of the feed quality for Phases 1 and 2 is presented in Table 42. Appendix Tables F-1 and F-2 present the results of all analyses performed on the RO feedwater. Approximately half way through Phase 2 of the test program the sampling point of the RO feedwater was changed from the DMF effluent

TABLE 41. REVERSE OSMOSIS MEMBRANE CHARACTERISTICS (16)

Type	TFC 4600 PA
Length	101.6 cm (40 in.)
Diameter	9.98 cm (3.93 in.)
Membrane material	Polyamide
Recommended initial operating pressure	281,000 - 361,000 kg/m ² (400 - 450 psi)
Maximum feedwater temperature	45°C
Recommended feedwater pH range	4.0 - 6.0
Initial product flow rate	4.73 m ³ /day (1,250 gpd)
% rejection based on chloride ion	97% minimum 98% average

tank to the pH adjustment tank just prior to the RO unit. This change was made so that the higher TDS concentration of the RO feedwater after pH adjustment could be determined. Addition of HCl to lower pH increases dissolved solids in the feedwater by adding chloride ions. This increase can be seen in the daily analyses presented in Appendix Table F-2. The total cyanide and cyanide amenable to chlorination concentrations were essentially equal throughout the study. The data indicate the pretreatment generally was adequate for operation of a spiral wound RO unit although the average turbidity of 2.4 FTU was slightly above the recommended value of 1.0 FTU. Feed turbidity above the recommended value indicates that more frequent cleaning of the membranes would be required and membrane life may be shortened.

The Millipore Pluggage Index of the dual media filter effluent (RO feed) was determined twice during the testing program. As a part of the bench testing phase, a pluggage index of 15 was observed. A grab sample of filter effluent taken during on-site testing had a pluggage index of 46. The nature of the test is such that indexes greater than 10 are essentially equal. A Millipore Pluggage Index of less than 3 is recommended for operation of a hollow fiber unit, although spiral wound membrane manufacturers generally use turbidity rather than pluggage index to determine waste treatability.

Cartridge filters of 10 micron size were provided with the unit to protect the high pressure pump and the membranes. The filters became plugged and required replacement every 33.7 hours on the average. The relatively long filter life is an indication of the lack of colloidal and suspended solids present in the system feedwater.

The membranes became fouled and required cleaning after a 47.6 hour run time in Phase I. The average feed turbidity for that run was 4.95 FTU. Run

TABLE 42. BLAST FURNACE REVERSE OSMOSIS FEED

Parameter	Minimum	Average	Maximum
pH	4.60	-	7.40
SS, mg/l	3	15 (12)	57
VSS, mg/l	1	4 (7)	6
TDS, mg/l	1119	1363 (12)	1648
Cond., μ mhos/cm	1500	2053 (11)	2350
Turb., FTU	0.50	2.43	25.0
Color, APHA units	0	1.4 (7)	5
TOC, mg/l	1	2 (11)	6
BOD ₅ , mg/l	<2	<2 (1)	<2
Phenol, mg/l	0.008	0.027 (11)	0.151
NH ₃ as N, mg/l	26.0	39.2 (12)	48.5
NO ₃ as N, mg/l	0.07	0.23 (6)	0.95
TKN as N, mg/l	26.4	38.0 (7)	50.2
S, mg/l	<0.2	<0.2 (7)	<0.2
CN, mg/l	<0.01	<0.05 (12)	0.32
CN _A , mg/l	<0.01	<0.05 (10)	0.27
F, mg/l	5.2	8.81 (10)	10.9
TP as P, mg/l	<0.01	<0.03 (6)	0.06
Cl, mg/l	380	419 (8)	520
SO ₄ , mg/l	155	183 (8)	220
Na, mg/l	16	18 (7)	27
Ca, mg/l	53	84 (7)	133
K, mg/l	22	24 (7)	26
Hg, mg/l	5.2	50 (7)	63
Mn, mg/l	2.6	3.6 (7)	5.2
Fe, mg/l	0.4	1.8 (7)	6.0
Alk., mg/l as CaCO ₃	250	280 (6)	330

() Indicates number of data points used to obtain average.

times during Phase 2 ranged from 2.2 - 172.5 hours. The run time, or fouling rate, was a function of the feed turbidity. When upsets occurred in the pretreatment processes, run times decreased significantly. A 4.0 hour run had an average turbidity of 4.25 FTU. The final 2 hours of that run had an average turbidity of 6.50 FTU. A 72.0 hour run had an average turbidity of 1.77 FTU, however the average turbidity for the last 3 hours of the run was 2.90 FTU. The 172.5 hour run had an average feed turbidity of 1.43 FTU. Again, poor quality feed apparently ended the run early. At approximately 160 hours into the run, the feedwater turbidity averaged 3.40 FTU for 9 hours. The Phase 2 test data indicate that all runs have a period of relatively poor quality feedwater that precipitated the end of the run. For this reason and assuming that a full-scale pretreatment system could avoid upsets and maintain an effluent turbidity of 1.5 FTU or less, the RO system will require cleaning every 6-7 days. This may be a conservative estimate of the fouling rate of a full-scale system.

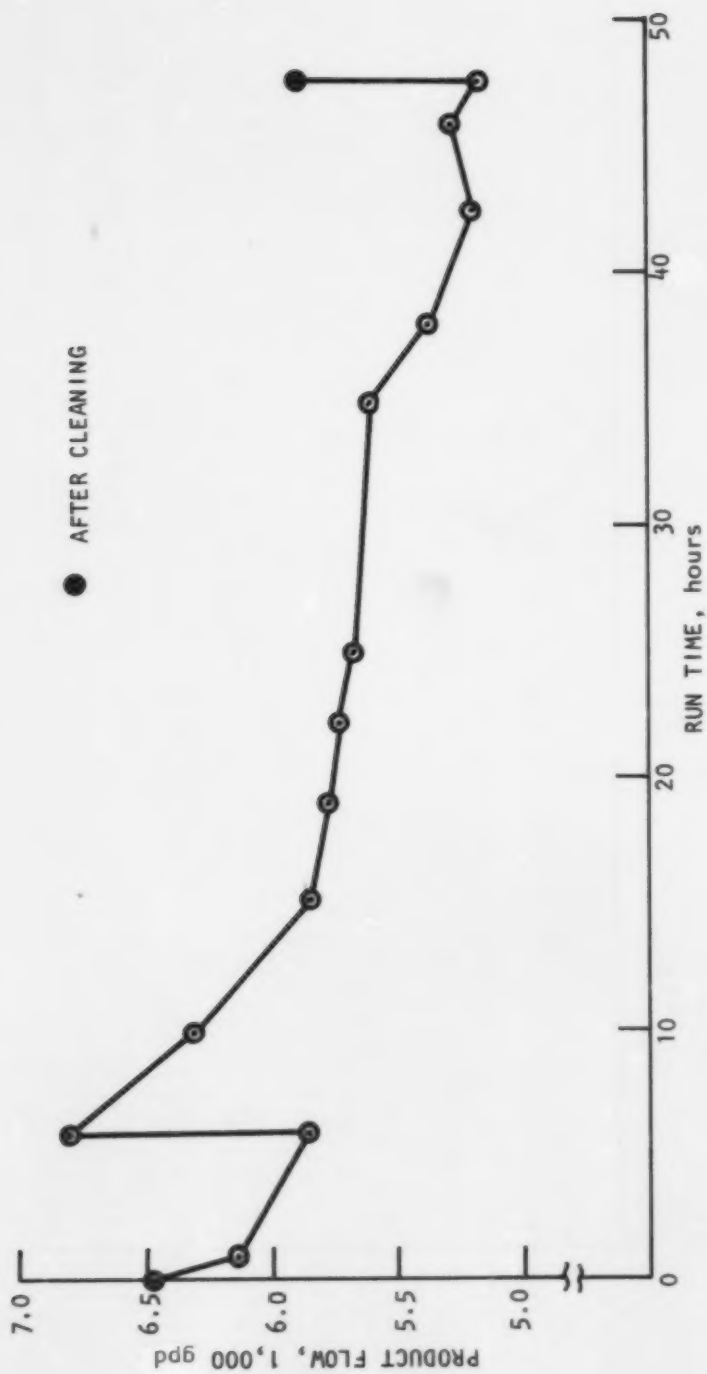
Feed turbidities were generally lower during Phase 2 than Phase 1, because lower loading rates could be utilized for the pretreatment processes. The use of brine recycle during Phase 2 lowers the effluent flow required from the pretreatment processes.

Flux Data

The RO system was operated 24 hours per day throughout the test program, except when mechanical or operational problems forced a shut down. The unit was operated as much as possible to better evaluate potential membrane life. A total of 409.5 hours, 47.6 hours during Phase 1 and 361.9 hours during Phase 2, were put on the unit.

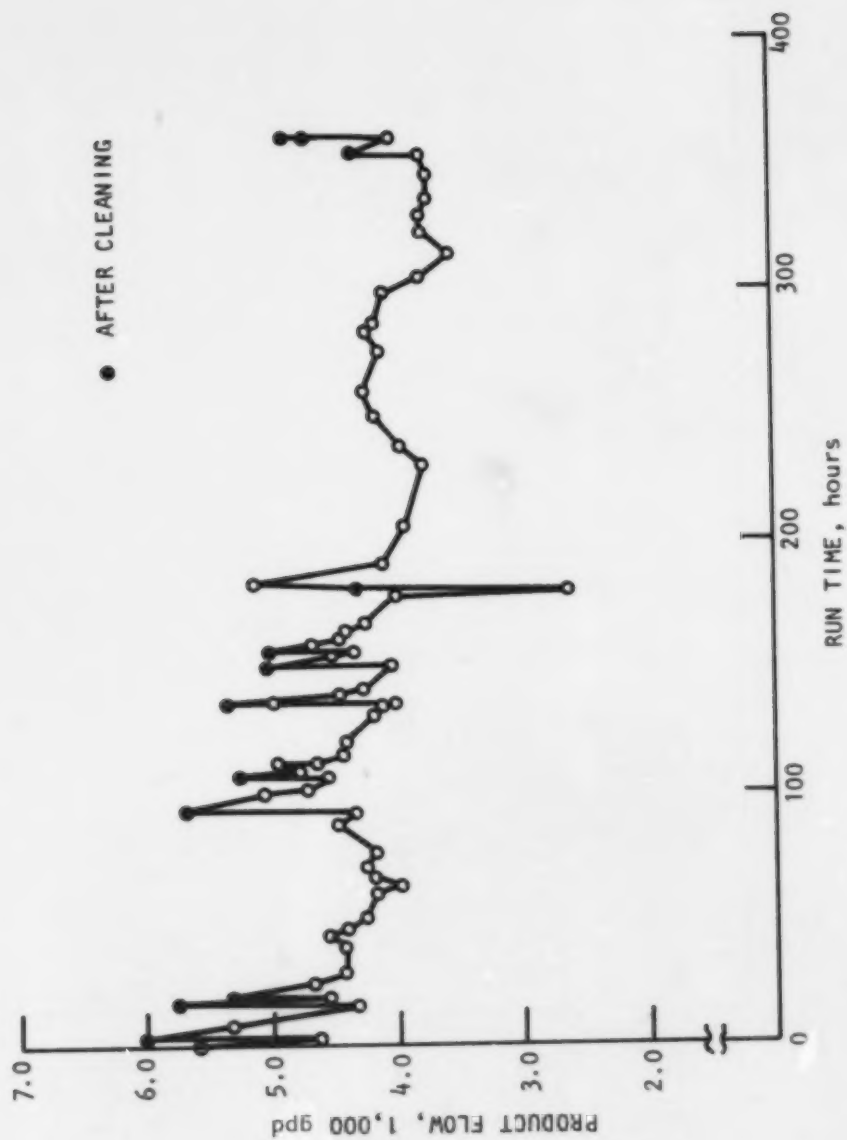
Plotted in Figure 53 is the product flow versus time data for Phase 1 of the test program. The data points have been corrected to assume operation at a feed pressure of 281,000 kg/m² (400 psig) to the first membrane and a feedwater temperature of 25°C; Appendix Table F-3 lists the corresponding operational data. The product water volumes represent the total produced by the three membranes. A flux decline (9.2 percent) over the first 50 hours can be seen. Since this decline can not be attributed to initial compaction of the membranes because of the PA membrane's resistance to compaction, it is believed that the fouling is attributable to the relatively high feedwater turbidity.

Figure 54 presents the product flow plotted versus time for Phase 2 of the test program. Again, the data points have been corrected to assume operation at 281,000 kg/m² (400 psig) feed pressure to the first membrane and a feedwater temperature of 25°C. Appendix Table F-3 lists the operational data this plot was based on. A steady flux decline of 26.7 percent occurred during Phase 2. Typically, reverse osmosis membrane life is considered over after a membrane flux decline of 30 percent. A double cleaning at the conclusion of Phase 2 restored the membrane flux 8.4 percent to 18.3 percent of the initial Phase 2 level. The decline is probably due to fouling from relatively high turbidity feedwater during pretreatment upsets. The two cleaning solutions used could not completely remove the fouling material.



NOTE: 1,000 gpd x 3.785 = m³/day

Figure 53. Product flow vs. run time - Phase 1.



NOTE: 1,000 gpd $\times$ 3.785 = m^3/day

Figure 54. Product flow vs. run time - Phase 2.

A total flux decline of 27.5 percent occurred during the test program. Phase 1 had a 9.2 percent decline and Phase 2 an 18.3 percent decline. The test program data indicate that a flux decline of 15-20 percent over the first 400-500 hours could be expected. If more efficient cleaning solutions could be found, the expected flux decline would be decreased.

Feed pressure and temperature during Phase 1 ranged from 152,000 - 179,000 kg/m² (215 - 255 psi) and 23.9 - 26.9 °C respectively. A feed pressure and temperature range of 149,000 - 281,000 kg/m² (212 - 400 psi) and 22.6 - 35.0 °C occurred during Phase 2.

Module Cleaning

Use of a citric acid cleaning solution and a soda ash (Na₂CO₃) cleaning solution was evaluated during Phase 2 of the test period. Fluid Systems recommends using a 2 percent citric acid (pH 3.0) solution and a 1.5 percent soda ash (pH 11.0) solution for cleaning their membranes (3). The citric acid solution is most effective in removing inorganic foulants while the soda ash solution is most effective in removing organic foulants.

The citric acid solution proved more effective than the soda ash solution and the soda ash solution was relatively ineffective alone. However, a double cleaning utilizing citric acid and soda ash was most effective. This indicates that membrane fouling is mainly inorganic in nature but that some organic fouling does occur. From the program results, regular cleaning with both citric acid and soda ash solutions is recommended for the full-scale system.

Normal cleaning periods were approximately one hour long. The cleaning process consisted of recirculating the solution through the membranes at pressures less than 42,200 kg/m² (60 psi) and flowrates of 8.3 - 17.0 l/min (2.2 - 4.5 gpm). After recirculation, the membranes were rinsed with feedwater for 15-30 minutes at less than 42,200 kg/m² (60 psi). The double cleaning consisted of recirculation of each solution for one hour and rinsing with feedwater for 15-30 minutes between solutions and after the final cleaning.

During the test program, the volume of cleaning solution amounted to 0.14 percent of the feed flow to the unit, based on a cleaning frequency of 7 days. The spent cleaning solutions will have a relatively high dissolved solids content. Table 43 presents an analysis of typical spent citric acid and soda ash cleaning solutions. The effectiveness of the citric acid solution when compared to the soda ash solution in removing inorganic foulants can be seen in Table 43. In most RO applications, the small volume of spent cleaning solution is returned to the head end of the treatment plant. At this particular site the spent cleaning solution would probably be sent to the mill's industrial wastewater treatment plant with minimal impact.

TABLE 43. REVERSE OSMOSIS SPENT CLEANING SOLUTIONS

Parameter	Citric acid ^a	Soda ash ^a
pH	2.40	11.30
SS, mg/l	16	92
VSS, mg/l	12	-
S ⁻ , mg/l	<0.2	-
F, mg/l	-	21
TOC, mg/l	-	<1
Color, APHA units		0/5
Cl, mg/l	210	-
SO ₄ , mg/l	90	-
Na, mg/l	23	-
K, mg/l	12	-
Ca, mg/l	40	3.5
Mg, mg/l	34	14
Mn, mg/l	1.9	0.05
Fe, mg/l	4.3	0.1

a. Analysis performed on one sample.

Mechanical and Operational Problems

The RO membranes exhibited a total flux decline of 27.5 percent during the test period. The useful life of RO membranes is normally considered over when the flux decline reached 30 percent. The program data indicates that a 15-20 percent flux decline over the first 400-500 hours could be expected from a full size system. This could be significantly reduced if pretreatment processes and cleaning methods were improved.

No significant mechanical problems occurred during the test program.

Effluent Quality

A summary of the average RO feed and product quality during Phase I of the test program is given in Table 44. Results of all analyses performed on the Phase I product water and feedwater are given in Appendix Tables F-1 and F-5. As can be seen, the rejection of all BAT parameters is quite good and the product water easily meets BAT discharge limitations. In a full-scale system, 25 percent of the total product volume would be of this quality.

TABLE 44. BLAST FURNACE REVERSE OSMOSIS PERFORMANCE - PHASE 1

Parameter	Feed ^a	Product ^a	Percent rejection
pH	7.60/7.70	5.35/5.65	
SS, mg/l	31	3	90
VSS, mg/l	4	1	75
Phenol, mg/l	0.013	0.115	-
NH ₃ as N, mg/l	34.9	5.2	85
TKN as N, mg/l	37.0	7.1	81
NO ₃ as N, mg/l	0.09	<0.01	>89
S, mg/l	<0.2	<0.2	-
CN, mg/l	0.06	<0.01	>83
F, mg/l	10.3	0.29	97.2
TOC, mg/l	3	<1	>67
TP, mg/l	<0.03	<0.01	>67
TDS, mg/l	1298	22	99
Cond., μ mhos/cm	1850	131	92.9
Color, APHA units	2	0	-
Alk., mg/l as CaCO ₃	280	17	94
Cl, mg/l	390	18	95
SO ₄ , mg/l	184	<2	>99
Na, mg/l	17	1.7	90
Ca, mg/l	56	2.1	96
K, mg/l	24	2.0	92
Mg, mg/l	56	2.1	96
Mn, mg/l	3.1	0.08	97
Fe, mg/l	3.2	0.6	81

a. Average obtained using three points.

Table 45 presents the average quality of the RO feed, product and brine during Phase 2. Appendix Tables F-2, F-6 and F-7 present the results of all analyses performed on the feedwater, product water, and brine during Phase 2. If discharged separately, this product water would not meet BAT limitations. However, when blended with the product water from the remainder of a full

TABLE 45. BLAST FURNACE REVERSE OSMOSIS PERFORMANCE - PHASE 2

Parameter	Feed	Product	Brine	Percent rejection
pH	6.50/7.90	5.50/8.20	6.50/7.50	
SS, mg/l	10 (12)	4 (4)	25 (3)	60
VSS, mg/l	3 (7)	3 (2)	- -	-
Phenol, mg/l	0.045 (10)	<0.054 (10)	0.276 (12)	-
NH ₃ as N, mg/l	36.0 (13)	20.0 (15)	115 (16)	44.4
NO ₃ as N, mg/l	0.37 (3)	<0.02 (4)	0.45 (6)	>94.6
TKN as N, mg/l	30.9 (6)	17.5 (6)	95.2 (6)	43.4
S, mg/l	<0.2 (6)	<0.2 (6)	<0.2 (7)	-
CN, mg/l	<0.04 (11)	<0.01 (13)	<0.15 (4)	>75
CN _A , mg/l	<0.04 (11)	<0.01 (13)	<0.15 (4)	>75
F, mg/l	7.50 (9)	0.63 (9)	18.0 (11)	91.6
TOC, mg/l	2 (12)	<2 (13)	14 (16)	-
BOD ₅ , mg/l	<2 (1)	<2 (1)	7 (1)	-
COD, mg/l	- -	5 (1)	61 (1)	-
TP as P, mg/l	<0.01 (3)	<0.01 (4)	<0.04 (4)	-
TDS, mg/l	1374 (13)	135 (15)	5375 (15)	90.2
Cond., μ mhos/cm	2110 (10)	410 (10)	7400 (12)	80.6
Color, APHA units	1 (4)	0 (5)	3.3 (6)	-
Alk., mg/l as CaCO ₃	280 (3)	43 (6)	520 (7)	85
Cl, mg/l	465 (7)	76 (9)	2180 (9)	84
SO ₄ , mg/l	190 (6)	<2 (5)	720 (3)	>99
Na, mg/l	19 (4)	4.8 (2)	75 (2)	75
Ca, mg/l	106 (4)	5.5 (2)	330 (2)	95
K, mg/l	24 (4)	6.2 (2)	580 (2)	74
Mg, mg/l	45 (4)	5.0 (2)	265 (2)	89
Mn, mg/l	4.1 (4)	0.29 (2)	21 (2)	93
Fe, mg/l	0.7 (4)	0.2 (2)	0.7 (2)	71

() Indicates number of points used to obtain average.

size system, BAT limitations would be met. Seventeen percent of the total product water produced in a full size unit would be similar in quality to this product water. The rejection of the various parameters are lower than those obtained during Phase 1, because of the use of the brine recycle loop to achieve 75 percent product water recovery. The parameter concentrations that the membranes actually see are significantly higher than those presented in Table 45. Sampling of the actual feed during recycle is possible with the pilot RO unit. Two grab samples of the actual feed were taken during the test program. A comparison of the feed plus brine recycle and the filter effluent is presented in Table 46.

A projection of the product water quality that could be expected from a full-scale RO system can be made from the data obtained in Phases 1 and 2. As can be seen in Table 47, reverse osmosis is capable of producing a high quality product water that meets BAT discharge limitations. It should be noted that NH_3 is the BAT parameter that is closest to exceeding the discharge limitations.

Side-stream Characteristics

A full size reverse osmosis system will produce three separate side-streams (brine, prefilter system backwash, and spent membrane cleaning solution) at various times during operation of the system. Of the three, brine will be the most important.

The RO system operating at 75 percent product water recovery will produce 3.785 liters (1 gallon) of brine for every 11.36 liters (3 gallons) of product water. The brine will contain most of the contaminants present in the feed at 3-4 times the original concentration because of the reduction in volume. The results presented in Table 45 are representative of brine that would be produced in a full size RO system. The brine contains high concentrations of dissolved salts, nutrients, and organics and would require further treatment to meet BAT discharge limitations.

A prefilter system consisting of a series of backwashable 10 micron size filters in parallel replaces the disposable cartridge filters used in the pilot study. Backwash from this system would be relatively small in volume, approximately 0.5 percent of the system feed flow. The backwash would have significant concentrations of suspended and colloidal solids but because of the small volume could be discharged to the plant industrial waste treatment plant with unknown impact.

As stated previously, the spent membrane cleaning solutions, amounting to 0.14 percent of the system feed flow, could also be sent to the industrial waste treatment plant.

Full Size System Design

Cost Information--

Cost estimates were prepared by an RO manufacturer for three full-scale designs, based on the test program results. The estimates were prepared

TABLE 46. COMPARISON OF DMF EFFLUENT AND
ACTUAL FEEDWATER DURING BRINE RECYCLE

Parameter	DMF Effluent	Actual Feed
pH	7.60-7.85	6.90-7.10
SS, mg/l	5 (2)	18 (1)
VSS, mg/l	3 (2)	13 (1)
Phenol, mg/l	0.019 (2)	0.015 (2)
NH ₃ as N, mg/l	46.0 (2)	87.0 (2)
NO ₃ as N, mg/l	0.10 (1)	0.18 (2)
TKN as N, mg/l	50.2 (1)	95.0 (1)
S, mg/l	<0.2 (1)	<0.2 (1)
CN, mg/l	<0.02 (1)	<0.02 (2)
CN _A , mg/l	<0.02 (1)	<0.02 (2)
F, mg/l	8.3 (2)	12.0 (2)
TOC, mg/l	2 (2)	4 (2)
TP as P, mg/l	<0.01 (2)	<0.01 (1)
TDS, mg/l	1420 (2)	3050 (2)
Cond., μ mhos/cm	2250 (2)	4200 (2)
Color, APHA units	0/5 (2)	0/5 (2)
Alk., mg/l as CaCO ₃	300 (2)	355 (1)
Cl, mg/l	413 (1)	1090 (2)
SO ₄ , mg/l	195 (1)	400 (1)
Na, mg/l	16 (2)	31 (1)
Ca, mg/l	86 (2)	300 (1)
K, mg/l	25 (2)	45 (1)
Mg, mg/l	56 (2)	130 (1)
Mn, mg/l	3.2 (2)	9.9 (1)
Fe, mg/l	0.9 (2)	0.4 (1)

() Indicates number of points used to obtain average.

TABLE 47. BLAST FURNACE PROJECTED REVERSE OSMOSIS PRODUCT QUALITY

Parameter	Product quality range
pH	5.35-5.75
SS, mg/l	3-4
VSS, mg/l	1-3
Phenol, mg/l	<0.054
NH ₃ as N, mg/l	9-10
NO ₃ as N, mg/l	<0.02
TKN as N, mg/l	10-12
S, mg/l	<0.2
CN, mg/l	<0.01
CN _A , mg/l	<0.01
F, mg/l	0.40-0.45
TOC, mg/l	<2
BOD ₅ , mg/l	<2
COD, mg/l	<5
TP as P, mg/l	<0.01
TDS, mg/l	60-70
Cond., μ mhos/cm	200-250
Color, APHA units	0
Alk., mg/l as CaCO ₃	25-30
Cl, mg/l	40-45
SO ₄ , mg/l	<2
Na, mg/l	2.5-3.5
Ca, mg/l	3.0-3.5
K, mg/l	3.5-4.0
Mg, mg/l	3.0-3.5
Mn, mg/l	0.15-0.20
Fe, mg/l	0.4-0.5

for system sizes of 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd), and 5,678 m³/day (1.5 mgd). Table 48 lists the system feedwater and operating characteristics used in preparing the cost information. In addition, the following cost guidelines were given:

1. Costs not to include pretreatment or brine disposal.
2. Costs not to include product water storage or distribution.
3. Manpower requirements not to include supervision or laboratory costs.
4. Reverse osmosis membranes assumed to have a 3 year life.

Listed in Table 49 are the cost estimates for the RO system. These costs are effective as of May, 1979. The polyamide membranes recommended by the manufacturer for this application were TFC Model 8600 PA membranes. The characteristics of this membrane are presented in Table 50.

The information in Table 49 is basically as received with the addition of costs for the low pressure pumps and the prefilter system. The low pressure pumps are required to pressurize the feed to 21,100 - 28,100 kg/m² (30 - 40 psi) prior to the high pressure pumps. The prefilter consists of a series of 10 micron size filters valved in parallel to common headers. The provision for backwash cleaning of each filter element in place, while the other elements remain on stream, is an important feature allowing continuous operation of the filter system. This system is more economical than a series of disposable cartridge filters. The backwash cleaning operation is illustrated in Figure 55. The filter element consists of a polypropylene sock and can be expected to have approximately a one year life. The 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd), and 5,678 m³/day (1.5 mgd) systems require one set of 3, 8, and 15 filter elements respectively.

Operating costs for the filter units are quite low. Filter element replacement once a year at \$13.50 per element would be \$202.50 per year for the largest system. Manpower requirements consist of one man backwashing the elements approximately three times daily. There are no power requirements.

The annual cost of a reverse osmosis system as a function of feed flow is presented in Figure 56. The information is presented to illustrate the economy of scale inherent in an RO system.

Space Requirements--

A process flow schematic provided by the manufacturer is shown in Figure 57. A decarbonator is provided to raise the permeate, or product water, pH from the 5.5 - 6.0 range to a more acceptable 6.0 - 8.0. Air is blown through the permeate, stripping dissolved CO₂ off as a gas and thereby raising the pH. Decarbonation would be required, however stabilization involving further pH control or chemical addition may also be necessary.

All of the items shown, with the possible exception of the decarbonator, would have to be enclosed within a building. Whenever possible, the filter system and the high and low pressure pumps should be placed in a separate

TABLE 48. REVERSE OSMOSIS FEED AND OPERATING CHARACTERISTICS

Feed Characteristics	Average	Range
Temperature, °C	-	22.8 - 35.0
pH	-	4.60 - 7.40
TDS, mg/l	1363	1119 - 1648
Conductivity, μ mhos/cm	2053	1500 - 2350
Turbidity, FTU	2.43	0.50 - 25.0
Color, APHA units	1.4	0 - 5
TOC, mg/l	<2	1 - 6
BOD ₅ , mg/l	2	0 - 2
Phenol, mg/l	0.027	0.008 - 0.151
NH ₃ as N, mg/l	39.2	26.0 - 48.5
NO ₃ as N, mg/l	0.23	0.07 - 0.95
TKN as N, mg/l	38.0	26.4 - 50.2
TP as P, mg/l	<0.03	<0.01 - 0.06
S ⁻ , mg/l	<0.2	0 - 0.2
CN, mg/l	<0.05	<0.01 - 0.32
CN _A , mg/l	<0.05	0.01 - 0.27
F, mg/l	8.81	5.2 - 10.9
Cl, mg/l	419	380 - 520
SO ₄ , mg/l	183	155 - 220
Na, mg/l	18	16 - 27
Ca, mg/l	84	53 - 133
K, mg/l	24	22 - 26
Mg, mg/l	50	5.2 - 63
Mn, mg/l	3.6	2.6 - 5.2
Fe, mg/l	1.8	0.4 - 6.0
Alkalinity, mg/l as CaCO ₃	280	250 - 330

Operating Characteristics:Feed pressure - 148,000-281,000 kg/m² (210-400 psi)

Product water recovery - 75%

Cleaning solutions - 2% citric acid, 1.5% Na₂CO₃ (Soda Ash)

TABLE 49. REVERSE OSMOSIS COSTS^a

System Size	946 m ³ /day (0.25 mgd)	2,839 m ³ /day (0.75 mgd)	5,678 m ³ /day (1.5 mgd)
Capital Cost^b			
RO system w/membrane	\$ 316,000	\$ 548,000	\$ 976,000
Feed pumps	5,000	9,000	14,000
Prefilters assembly	3,000	8,000	13,000
Delivered Equipment Total	\$ 324,000	\$ 565,000	\$1,003,000
Equipment installation	130,000	226,000	401,000
Installed Equipment Total	\$ 454,000	\$ 791,000	\$1,404,000
Process piping	136,000	237,000	421,000
Instrumentation	23,000	40,000	70,000
Capital Equipment Total	\$ 613,000	\$1,068,000	\$1,895,000
Facilities & Fees	1,318,000	2,297,000	4,075,000
Total Plant Cost	\$1,931,000	\$3,365,000	\$5,970,000
Operating Costs, \$/Year			
Operator ^c	\$ 54,000	\$ 54,000	\$ 54,000
Cleaning chemicals	3,000	10,000	20,000
Other chemicals ^d	9,000	27,000	53,000
Membrane replacement ^e	21,000	49,000	99,000
Electrical ^b	23,000	68,000	136,000
Maintenance labor & materials	31,000	53,000	95,000
Subtotal	\$ 141,000	\$ 261,000	\$ 457,000
Amortization of Capital ^g	181,000	315,000	558,000
Total Annual \$/Year	\$ 322,000	\$ 576,000	\$1,015,000
\$/3,785 L (\$/1,000 gal.) ^h	\$3.53	\$2.10	\$1.85

a. Excluding pretreatment and brine disposal.

b. See Appendix Tables A-4 and A-5 for basis of estimate.

c. Skilled labor, \$18,000/year.

d. Acid and scale inhibitor.

e. 3 year membrane life.

f. \$0.03/KWH.

g. 20 year service life, 6.875%.

h. Feed water.

TABLE 50. TFC 8600 PA MEMBRANE CHARACTERISTICS

Length	101.6 cm (40 in.)
Diameter	20.0 cm (7.88 in.)
Membrane material	polyamide
Recommended initial operating pressure	281,000 - 316,000 kg/m ² (400 - 450 psi)
Maximum feedwater temperature	45 °C
Recommended feedwater pH range	4.0 - 6.0
Initial product flow rate	18.9 m ³ /day (5,000 gpd)
% Rejection based on chloride ion	97% minimum 98% average

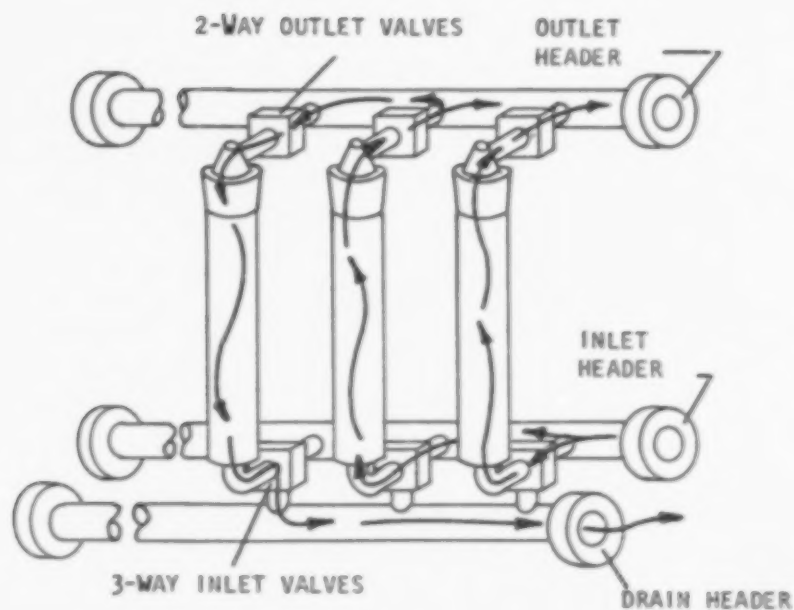
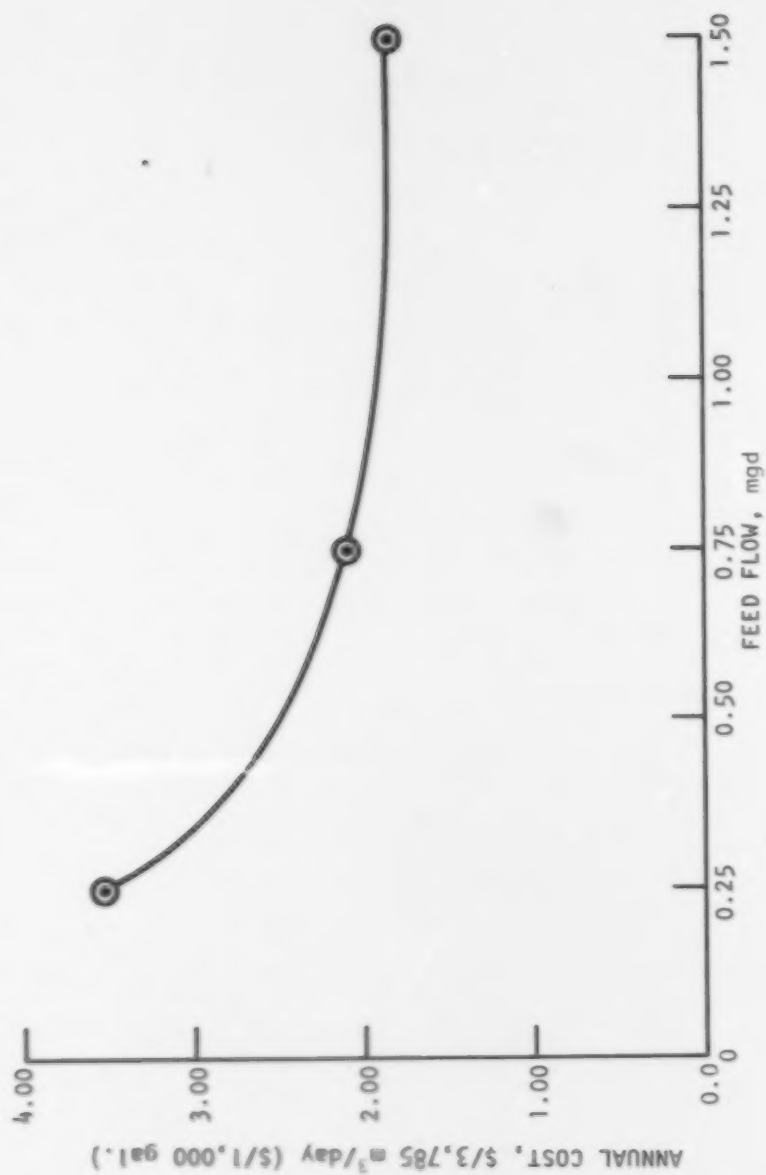


Figure 55. Schematic showing backwash cleaning of filter element.



NOTE: $\text{mgd} \times 3,785 = \text{m}^3/\text{day}$

Figure 56. Reverse osmosis system annual cost.

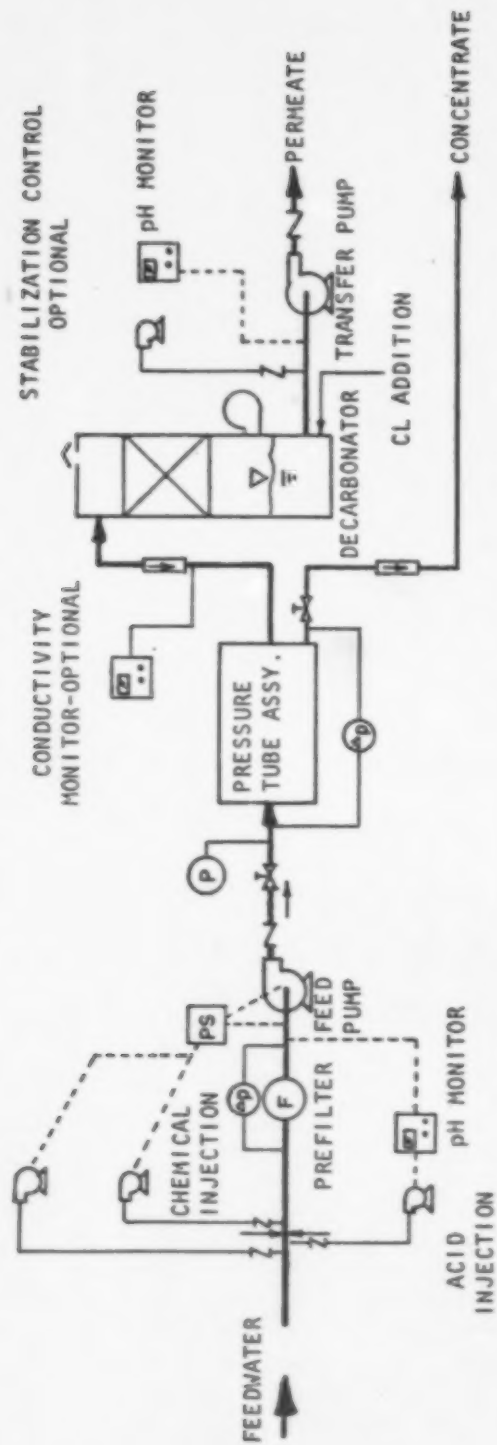


Figure 57. Reverse osmosis process flow schematic.

building or compartment. A significant reduction in the noise level normally associated with an RO installation would be obtained by isolation of the pumps in a well insulated area. The estimated space requirements for the RO systems are given in Table 51.

TABLE 51. REVERSE OSMOSIS SYSTEM SPACE REQUIREMENTS

System size	Area	
	Pump building	Remaining equipment
946 m ³ /day (0.25 mgd)	9.3 m ² (100 ft ²)	62 m ² (668 ft ²)
2839 m ³ /day (0.75 mgd)	21.6 m ² (233 ft ²)	145 m ² (1560 ft ²)
5678 m ³ /day (1.5 mgd)	43.3 m ² (466 ft ²)	290 m ² (3120 ft ²)

RESULTS OF THE OZONATION STUDY

General

The test program investigated the efficiency of ozonation in meeting BAT discharge limitations as a component of two treatment trains. The first train consisted of dual media filtration of the raw wastewater with ozonation and clarification of the filter effluent. The second train consisted of chemical treatment, clarification, filtration, reverse osmosis (RO) treatment of the raw wastewater, and ozonation of the RO brine. Operational variables and their effect on treatment efficiency investigated in the test program were hydraulic retention time (contact time), mass transfer, and ozone dosage. The objective of the test program was to determine if BAT discharge limitations could be met utilizing ozonation. The bench scale testing indicated that the ammonia limitation was the critical factor in meeting the discharge limitations. For this reason, the ozonation studies were geared to attempting to meet the ammonia limitation.

Filtered Wastewater Ozonation

The filtered wastewater ozonation portion of the test program was performed from January 9-22, 1979. The treatment train consisted of dual media filtration, pH adjustment, ozonation, neutralization, and clarification of the raw wastewater. Table 52 presents the raw wastewater quality during the ozonation study. The results of all analyses performed on the raw wastewater during the ozonation study are also presented in Table 53. The raw wastewater quality is presented because the dual media filter (DMF) would essentially remove only the suspended solids in the wastewater and because a more complete analysis of the raw water quality is available.

After filtration, the pH of the wastewater was raised from the normal 7.10-7.95 range to 10.5-11.5 range with 2.5 percent NaOH. This pH elevation was

TABLE 52. BLAST FURNACE RAW WASTEWATER QUALITY

Parameter	Average	Range
pH		7.10 - 7.95
SS, mg/l	28 (5)	4 - 40
VSS, mg/l	11 (1)	11
Phenol, mg/l	0.221 (9)	0.025 - 1.10
NH ₃ as N, mg/l	46.0 (11)	29.9 - 72.1
NO ₃ as N, mg/l	0.12 (3)	0.03 - 0.16
TKN as N, mg/l	46.9 (9)	29.6 - 64.5
S, mg/l	<0.2 (8)	<0.1 - 0.3
CN, mg/l	<0.44 (6)	<0.01 - 0.57
CN _A , mg/l	<0.29 (9)	<0.01 - 0.57
F, mg/l	12.6 (9)	10.0 - 17.2
TOC, mg/l	24 (9)	10 - 60
BOD ₅ , mg/l	18 (2)	11 - 24
Color, APHA units		5/10
O&G, mg/l	<4 (2)	<1 - 6

() Indicates number of points used to obtain average.

necessary to make the ammonia more available for oxidation by the ozone. A pH controller was used to maintain the desired range. Following ozonation, the wastewater was neutralized using 1.0 percent HCl. Polyelectrolyte addition and clarification completed the treatment train.

The ranges of operating data for this phase of the testing were as follows:

Total ozone dosage applied 0-1500 mg/l
 Total ozone dosage utilized 0-612 mg/l
 Influent wastewater flow rate 4.73-18.9 l/min (1.25-5.0 gpm)
 Hydraulic contact time 60-240 minutes
 Gas flow rate 164-510 l/min (5.8-18.0 scfm)

The operating data these ranges were taken from is presented in Appendix Table G-1.

Table 54 presents a summary of the effluent quality for each run during the study. The first series of test runs in the study incorporated a 60 minute contact time and various O₃ dosages in an attempt to reach an effluent NH₃

TABLE 53. BLAST FURNACE RAW WASTEWATER QUALITY

Parameter	DATE									
	1/10	1/11	1/12	1/13	1/15	1/16	1/17	1/19	1/20	1/22
pH	7.60		7.95	7.10	7.10	7.20		7.50	7.70	7.10
SS, mg/l	33		4	29	33	40				7.10
VSS, mg/l	11									
Phenol, mg/l	0.040			0.038	1.10	0.610	0.034	0.038	0.025	0.060
NH ₃ as N, mg/l	46	55	72.1	58.6	42.9	41.4	43.9	37.7	29.9	38
NO ₃ as N, mg/l						0.11	0.08	0.16		
TKN as N, mg/l	53	60		64.5	47.0		44	39.0	29.6	41.3
S, mg/l	<0.2			<0.2	<0.2	0.3	<0.2	0.3		<0.1
CN, mg/l						0.35	0.57	<0.01	0.33	0.49
CN _A , mg/l	0.30			<0.01	0.25	0.35	0.57	<0.01	0.33	0.49
F, mg/l	12.2			11.2	13.5	11	10.5	10	12.5	17.2
TOC, mg/l	12			10	14	60	18	34	38	20
BOD ₅ , mg/l						24	11.1			
Color, APHA units				5/10	5	5	5	5	5	5
O & G, mg/l				<1		6				

TABLE 54. BLAST FURNACE OZONATED WASTEWATER QUALITY

Parameter	DATE											
	1/10	1/11	1/12	1/13	1/15	1/16	1/17	1/19	1/19	1/20	1/21	1/22
Run number	1	2	4	6	7	8	10	12	13	14	15	16
Contact time, min.	60	60	60	240	240	240	120	120	120	171	120	120
O ₃ applied, mg/l	269	323	1500	1054	460	695	1205	735	781	1074	0	739
O ₃ utilized, mg/l	176	286		363	359	275	364	329	410	480	0	612
Parameter												
pH				7.90	7.80	7.40		7.60		6.90	7.50	7.60
SS, mg/l			16	1	3	9	7	5	8	15	7	16
VSS, mg/l				<1	<3	2	3	2	3	5	1	5
Phenol, mg/l	0.204			0.010	0.006	<0.001		0.002	0.176	0.005	0.660	0.190
NH ₃ as N, mg/l	34	46	40.5	3.1	29.0	13.7	9.4	13.0	11.6	3.4	32	10
NO ₃ as N, mg/l						25.5	34.5	38.5				
TKN as N, mg/l		52		1.1	31.3		4.8	11.0	3.5	0.67	35.2	12.7
S, mg/l				<0.2	<0.2		<0.2	0.3	<0.2	<0.2	<0.1	0.1
CN, mg/l				<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.55	0.03
CN _A , mg/l	0.02			<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.50	0.03
F, mg/l				10.8	15.3	11.5	11	10.5	11	12	15.6	16.2
TOC, mg/l	6			<2	<2	3	6	4	4	5	<1	<1
BOD, mg/l						5.9	<5		6.9			
Color, APHA units			5	5/10	10	0/5	0/5	5	10	5	10	10
O & G, mg/l				<1		8						

concentration of 10 mg/l or less. The applied ozone concentrations ranged from 269-1500 mg/l. The effluent NH_3 concentrations ranged from 34-46 mg/l, corresponding to percent removals of 16-26 percent. The presence of unreacted ozone in the contactor off-gases in all runs indicated that not all the ozone applied was being utilized. The mass transfer efficiency of the pilot units was not high enough to provide sufficient ozone for oxidation of the ammonia. For this reason, longer contact times were required to provide sufficient available ozone for oxidation of the ammonia. The pH of the ozonated wastewater before neutralization was below the desired 10.5-11.5 range for the first several runs. Oxidation products present in the ozonated wastewater lowered the pH to a 9.0-9.5 range. Several workers (17,18,19) have found that oxidation with ozone of solutions containing ammonia, phenol, and cyanide results in a lowering of the solution pH. The reaction products of ammonia oxidation with ozone are said to be nitrate ion (NO_3^-), molecular oxygen (O_2), water (H_2O), and hydrogen ion (H^+). Complete oxidation of phenol with ozone is said to produce carbon dioxide (CO_2), water (H_2O), and molecular oxygen (O_2). The oxidation of cyanide involves two separate chemical steps, oxidation of cyanide to cyanate (CNO^-) and oxidation of cyanate to nitrogen and ammonia. It was determined that a final pH range less than 10.5-11.0 would decrease the susceptibility of ammonia to oxidation. For this reason, all remaining test runs utilized an initial pH range that would produce a final pH range of 10.5-11.0. This initial pH range typically was 11.5-12.0.

The remaining test runs of the study investigated the effect of contact time and O_3 dose on treatment efficiency. The first run of this series used a 120 minute contact time and a 500 mg/l applied O_3 dose. The clarified effluent still had an ammonia concentration of 33 mg/l. The next three runs used a 240 minute contact time in conjunction with applied O_3 dosages of 1054, 460, and 695 mg/l. The 1054 mg/l ozone dose proved quite effective, producing a final effluent ammonia concentration of 3.1 mg/l. The amount of ozone actually utilized in the oxidation was determined to be 363 mg/l. Because of the relatively low percentage of ozone actually utilized, an applied ozone dose of 460 mg/l was used in the next test run. This run produced an effluent ammonia concentration of 29.0 mg/l. The amount of ozone utilized was 359 mg/l, nearly identical to the previous run. It was determined that the higher effluent NH_3 concentration was due to the above normal concentration of phenol (1.10 mg/l compared to an average of 0.204 mg/l) in the ozonation system feedwater. Apparently, the increased ozone demand of the phenol reduced the amount of ozone available to oxidize NH_3 . Therefore, it is believed that ozone preferentially attacks phenol over ammonia in the wastewater. The following run used an applied ozone dose of 695 mg/l. An effluent ammonia concentration of 13.7 mg/l was achieved in this run. A relatively low 275 mg/l of ozone was utilized in this run. The phenol concentration of the feedwater was still above average at 0.610 mg/l.

Four runs were conducted using 120 minute contact times and higher applied ozone dosages. Higher applied dosages were utilized to investigate whether higher available concentrations of ozone would increase the mass transfer driving force and therefore increase the amount of ozone utilized. The first run, using a 1000 mg/l applied dose, produced an effluent ammonia concentration of 21 mg/l. Data were not available to determine ozone utilized for

this run. The second run, using a 1205 mg/l applied dose, produced an effluent ammonia concentration of 9.4 mg/l. The system feedwater during this run contained the highest cyanide concentration (0.57 mg/l compared to an average of <0.44 mg/l) observed during the study period. It should be noted at this time that throughout the study, the total cyanide and cyanide amenable to chlorination concentrations were essentially equal. It is not apparent what effect, if any, the higher concentration had on the ozone demand and effluent ammonia. The concentration of ozone utilized in this run was 364 mg/l. The final two runs of this series, using a 735 and 781 mg/l applied doses respectively, produced final effluent NH_3 concentrations of 13.0 and 11.6 mg/l. These two runs were performed on the same day, therefore any difference in effluent concentration can be attributed to ozone dose only, because the feedwater characteristics are essentially the same. The improvement in effluent concentration from 13.0 to 11.6 mg/l coincides with an increase in ozone utilized from 329 to 410 mg/l. This indicates that under similar conditions, an increase in available ozone increases the amount utilized and thereby decreases the effluent ammonia concentration. However, in general, the four runs at higher dosages did not utilize significantly increased amounts of ozone than the previous test runs. This may be due to the feedwater quality variations discussed earlier.

The next test run in the program investigated a 171 minute contact time and an applied O_3 dose of 1074 mg/l. The 171 minute contact time was selected to provide a contact time between the 120 and 240 minutes previously used. An effluent concentration of 3.4 mg/l was produced while utilizing 480 mg/l of ozone.

The final two runs of the study period used 120 minute contact times. The first run was conducted under the same conditions as the previous runs except that the ozone generator was not turned on. This run demonstrated that the ammonia removal occurring in the test runs was due to oxidation by ozone and not air stripping of the ammonia gas at an elevated pH. The system feedwater and effluent NH_3 concentrations were 38 and 32 mg/l for this run, respectively. The final run, using an applied dose of 739 mg/l, produced an effluent of 10 mg/l NH_3 with 612 mg/l of ozone utilized. The reason for the high quantity of ozone utilized in this reaction is unknown. The concentrations of parameters in the feedwater are generally below the average for the study period.

Further confirmation that the ammonia removed was oxidized by the ozone was obtained from three runs during the study. Previous work (17) has shown that nitrate is the principal end-product of the oxidation of ammonia with ozone. As part of the three test runs conducted on January 16, 17, and 19, the system influent and effluent nitrate concentrations were analyzed along with the normal parameters. The change in influent and effluent ammonia and nitrate concentrations should be approximately equal if oxidation of ammonia does occur. Table 55 presents the results of these tests. The data indicate that there is a good correlation between the reduction in ammonia and the increase in nitrate concentrations. The reason for the discrepancy between ammonia reduction and nitrate increase on January 19 cannot be determined from the data. The additional nitrate may have been the result

TABLE 55. PRODUCTION OF NITRATE DURING OZONATION

	Date		
	1/16	1/17	1/19
NH ₃ as N, mg/l			
Influent	41.4	43.9	37.7
Effluent	13.7	9.4	13.0
NH ₃ Reduction, mg/l	27.7	34.5	24.7
NO ₃ as N, mg/l			
Influent	0.11	0.08	0.16
Effluent	25.5	34.5	38.5
NO ₃ Production, mg/l	25.39	34.42	28.34
NH ₃ + NO ₃ as N, mg/l			
Influent	41.51	43.98	37.86
Effluent	39.2	43.9	51.5

of other unknown oxidation products or a sampling or analytical error.

Further investigation into the effect of contact time on treatment efficiency was conducted on two separate runs. The ozonation system permitted sampling of the wastewater between the three 378.5 liters (100 gallon) contactors. Figure 58 presents a graph of the wastewater ammonia concentration as a function of contact time for two runs. The runs utilized total contact time of 120 minutes, 40 minutes per contactor. Figure 58 demonstrated that the reduction in ammonia did occur throughout the total contact time.

The pilot scale data demonstrated that BAT discharge limitations could be met by a treatment train consisting of filtration, ozonation, and clarification. Run numbers 6, 10, 14 and 16 met all BAT parameters. Three different contact times, ranging from 120-240 minutes, were used in these four runs. The concentrations of ozone utilized ranged from 363-612 mg/l. Table 56 presents the effluent quality and important operational parameters for the four runs. Run numbers 6 and 10 confirm the assumption that shorter contact times can achieve results comparable to longer contact times if ozone mass transfer can be increased sufficiently by using higher applied dosages. This indicates that ozone mass transfer is the controlling factor in the oxidation of ammonia.

The ratios of ozone required/NH₃ removed for the runs given in Table 56 are

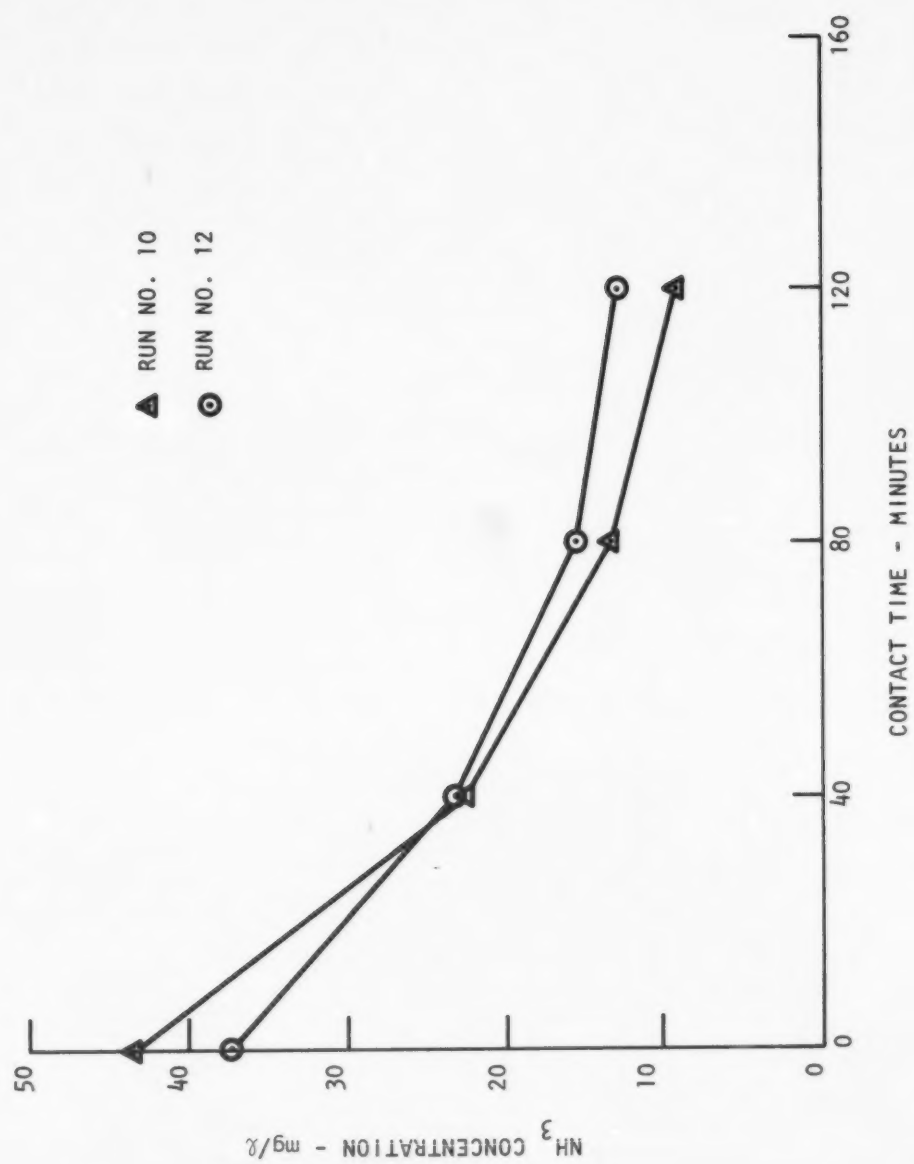


Figure 58. Effluent NH_3 vs. contact time.

TABLE 56. OZONATION RUNS MEETING BAT LIMITATIONS

	Date			
	1/13	1/17	1/20	1/22
Run number	6	10	14	16
Contact time, min.	240	120	171	120
O ₃ applied, mg/l	1054	1205	1074	739
O ₃ utilized, mg/l	363	364	480	612
Influent NH ₃ as N, mg/l	58.6	43.9	29.9	41.0
pH	7.90		6.90	7.60
SS, mg/l	1	7	15	16
VSS, mg/l	<1	3	5	5
Phenol, mg/l	0.010		0.005	0.190
NH ₃ as N, mg/l	3.1	9.4	3.4	10
NO ₃ as N, mg/l		34.5		
TKN as N, mg/l	1.1	4.8	0.67	12.7
S, mg/l	<0.2	<0.2	<0.2	0.1
CN, mg/l		<0.01	<0.01	0.03
CN _A , mg/l	<0.01	<0.01	<0.01	0.03
F, mg/l	10.8	11.0	12.0	16.2
TOC, mg/l	<2	6	5	<1
BOD ₅ , mg/l		<5		
Color, APHA units	5/10	0/5	5	10
O&G, mg/l	<1			
O ₃ /NH ₃ , mg/mg	6.54	10.5	18.1	19.7

quite variable. This is probably due to the fact that the amount of ozone required is dependent on the concentrations of other parameters along with ammonia. The variation in system feedwater quality is reflected in these ratios. The bench tests had indicated that 23.1 mg O₃ would be required to remove 1 mg of NH₃.

Reverse Osmosis Brine Ozonation

The RO brine ozonation portion of the test program was performed from March 7-15, 1979. The treatment train consisted of chemical treatment, clarifica-

tion, dual media filtration, reverse osmosis treatment of the raw wastewater, and ozonation of the concentrated reverse osmosis brine. Table 57 presents a summary of the RO brine quality during the ozonation study. The results of all analyses performed on the brine during the study are presented in Table 58. As found in the filtered wastewater ozonation study, the concentrations of total cyanide and cyanide amenable to chlorination were essentially equal over the study period. The pH of the brine was raised from the normal 6.4 - 6.9 range to 11.5 - 12.0 with 2.5 percent NaOH to increase the availability of ammonia for oxidation. Following ozonation, the wastewater was neutralized using 1.0 percent HCl. Polyelectrolyte addition and clarification completed the treatment train.

TABLE 57. BLAST FURNACE REVERSE OSMOSIS BRINE - OZONATION FEED

Parameter	Minimum	Average	Maximum
pH	7.10	-	7.50
SS, mg/l	21	24 (2)	26
Phenol, mg/l	0.029	0.390 (5)	0.595
NH ₃ as N, mg/l	71	125 (10)	172
NO ₃ as N, mg/l	0.27	0.47 (4)	0.60
CN, mg/l	<0.01	<0.24 (8)	0.95
CN _A , mg/l	<0.01	<0.24 (8)	0.95
S, mg/l	<0.2	<0.2 (2)	<0.2
F, mg/l	6.6	17.5 (5)	25.1
TOC, mg/l	12	15.8 (8)	24
TP, mg/l	0.01	0.01 (2)	<0.02
TDS, mg/l	4540	5131 (7)	6119
Cl, mg/l	1800	2085 (3)	2325
SO ₄ , mg/l	675	675 (1)	675
Color, APHA units	0	4 (4)	5
Alk. as CaCO ₃ , mg/l	588	591 (2)	593
Cond., μ mhos/cm	6300	7300 (6)	7800
BOD ₅ , mg/l	7	7 (1)	7

() Indicates number of points used to obtain average.

The ranges of operating data for this phase of the testing were as follows:

Total ozone dosage applied 0 - 3,065 mg/l
 Total ozone dosage utilized 0 - 1,465 mg/l
 Influent wastewater flowrate 2.84 - 9.46 L/min (0.75 - 2.50 gpm)
 Hydraulic contact time 120 - 400 min.
 Gas flow rate 208 - 462 L/min (7.34 - 16.3 scfm)

The operating data these ranges were taken from are presented in Appendix Table G-2.

TABLE 58. BLAST FURNACE REVERSE OSMOSIS - OZONATION FEED

Parameter	DATE									
	3/7	3/8	3/9	3/10	3/11	3/12	3/13	3/14	3/15	3/16
pH	7.20	7.20	7.30		7.30	7.10	7.20	7.50	7.50	
SS, mg/L		36						21		
Phenol, mg/L	0.535	0.563				0.595	0.029		0.228	
NH ₃ as N, mg/L	130.5	117	136	118 ^a	100	159	115	133	172	71
NO ₃ as N, mg/L		0.27	0.55			0.60		0.44		
TKN as N, mg/L								116.5		
TP as P, mg/L		0.01						<0.02	0.38	
CN, mg/L	0.04	<0.01	<0.01		<0.01	<0.01	0.51	0.95	0.38	
CN _A , mg/L	0.01	<0.01	<0.01		<0.01	<0.01	0.51	0.95	21.3	
S, mg/L	<0.2						<0.2		14	
F, mg/L			13.0			6.6	21.7	25.1	5	
TOC, mg/L	13	12	13		17	13	20	24	6300	
BOD ₅ , mg/L		7								
Color, APHA units	5	0/5						5		
Cond., μ mhos/cm	8500	6700			7000		7500	7800	4665	
Alk., mg/L as CaCO ₃		593						588	1800	
TDS, mg/L	6119	4794			4840	4540	5262	5695		
Cl, mg/L	2325							2130		
SO ₄ , mg/L		675								
Na, mg/L		50						100		
Ca, mg/L		560						94		
K, mg/L		75						1090		
Mg, mg/L		240						290		
Mn, mg/L		17						24		
Fe, mg/L		0.8						0.5		

^a On-site measurement

A summary of the effluent quality and important operating parameters for each run is presented in Table 59. Run numbers 1 and 2 indicated that longer contact times and higher applied dosages would be required. As in the filtered wastewater ozonation study, the mass transfer efficiency of the pilot unit was not high enough to supply sufficient ozone for the reaction. Shorter contact times and lower ozone dosages could be utilized with units capable of more efficient mass transfer. Run numbers 3 and 5, respectively, showed some improvement in ammonia oxidation. Run number 4 was conducted under the same conditions as runs 3 and 5 except that the ozone generator was not turned on. The influent and effluent ammonia concentrations of 118 and 95 mg/l indicate that the majority of the ammonia removal is by oxidation and not air stripping at an elevated pH. A 300 minute contact time was used in runs 6 and 7. Run 7, with an applied ozone dose of 1,704 mg/l, used 665 mg/l of ozone to reduce the effluent ammonia concentration to 42 mg/l. Runs 8 and 9 used a 400 minute contact time and high applied ozone dosages of 1,877 and 3,000 mg/l in an effort to meet BAT discharge limitations. Run 8 produced an effluent ammonia concentration of 42 mg/l, the same as Run 7, even though more ozone was utilized, 747 vs 665 mg/l. This can be explained by comparing the brine quality for those runs. The brine encountered during Run 8 had a higher ammonia concentration, 133 vs. 115 mg/l, and cyanide concentration, 0.95 vs. 0.51 mg/l than did Run 7. The poorer quality brine required more ozone and a longer contact time. Run 9 used 1,465 mg/l ozone to produce an effluent ammonia concentration of 30 mg/l. This is the highest concentration of ozone utilized in the study and also the lowest effluent ammonia concentration. Unfortunately, the brine ammonia concentration of 172 mg/l was also the highest of the study. Even with a 142 mg/l ammonia reduction, the BAT ammonia discharge limitation was not met. A reduction in ammonia concentration of 140 mg/l would meet BAT ammonia limitations for all but one of the test runs. Run 10 was a final attempt to reduce the ammonia concentration to an acceptable level. A batch run using pH adjusted reverse osmosis brine and one 378.5 liters (100 gal.) contactor was performed. The ammonia concentration was reduced to 10 mg/l after 180 minutes using a total applied dose of 3,065 mg/l. Data needed to determine the amount of ozone utilized could not be obtained during the run due to an operating problem with the system. Unfortunately, the ammonia concentration of the RO brine used in the batch run was only 71 mg/l. The low initial ammonia concentration made the special run and removal efficiency atypical.

Figure 59 shows the variation in RO brine ammonia concentrations during the study period. The large variation made the daily planning necessary during the study difficult since contact times and applied ozone dosages were based on the results of the previous day's run. Decisions were made assuming a relatively constant RO brine concentration from day to day. This wide variation was also seen in the ammonia concentrations of the reverse osmosis product water indicating that the variation was present in the raw wastewater.

The average ammonia concentration for the study of 125 mg/l would require 87 percent removal to meet the BAT discharge limit.

TABLE 59. BLAST FURNACE OZONATED REVERSE OSMOSIS BRINE EFFLUENT

Parameter	DATE											
	3/7	3/8	3/9	3/10	3/11	3/12	3/13	3/14	3/15	3/16		
Run number	1	2	3	4	5	6	7	8	9	10		
Contact time, min.	120	120	240	240	240	300	300	400	400	180		
O ₃ applied, mg/l	991	738	1176	0	1319	632	1704	1877	3000	3065		
O ₃ utilized, mg/l	760	470	537	0	525	169	665	747	1465	a		
pH	6.90	7.80	7.80	5.90	6.60	8.40	6.30	9.90	8.50			
SS, mg/l	38	31	37	76	39	74	78	53	65			
Phenol, mg/l	<0.001	0.002	0.018	0.012		0.012	0.007	0.023	0.56			
NH ₃ as N, mg/l	39	91.5	75	95	57	69	42	42	30 ^b	10		
NO ₃ as N, mg/l		32.0	48	15		33	73	85	78			
TP as P, mg/l								0.02				
CN _A , mg/l	<0.01	<0.01	<0.01		<0.01	<0.01	<0.01	<0.01	<0.01			
CN, mg/l	<0.01	<0.01	<0.01		<0.01	<0.01	<0.01	<0.01	<0.01			
F, mg/l	17.4	19.3	16.3	25	22	18.7	27.6	9.1	14.0			
TOC, mg/l	22	7		7	13	7	4	6	4			
BOD ₅ , mg/l		9										
Color, APHA units	0				0/5	0/5		0	0			
Cond., μ hos/cm	14500	14500					13500	13000	14500			
Alk., mg/l as CaCO ₃		245										
TDS, mg/l	9795	8669	8073	8660	7750	4740	10626	8353	9652			
Cl, mg/l	4925			4300		3500	5600	4230	4850			
SO ₄ , mg/l		725										
Contactor No. 1 effluent												
NH ₃ as N, mg/l	41	117	65		46	64	42	55	66			
NO ₃ as N, mg/l		15	250									
Contactor No. 2 effluent												
NH ₃ as N, mg/l	35	85.5	64		44	61	36	41	70			
NO ₃ as N, mg/l		55	240									

a. Data not available

b. On-site measurement

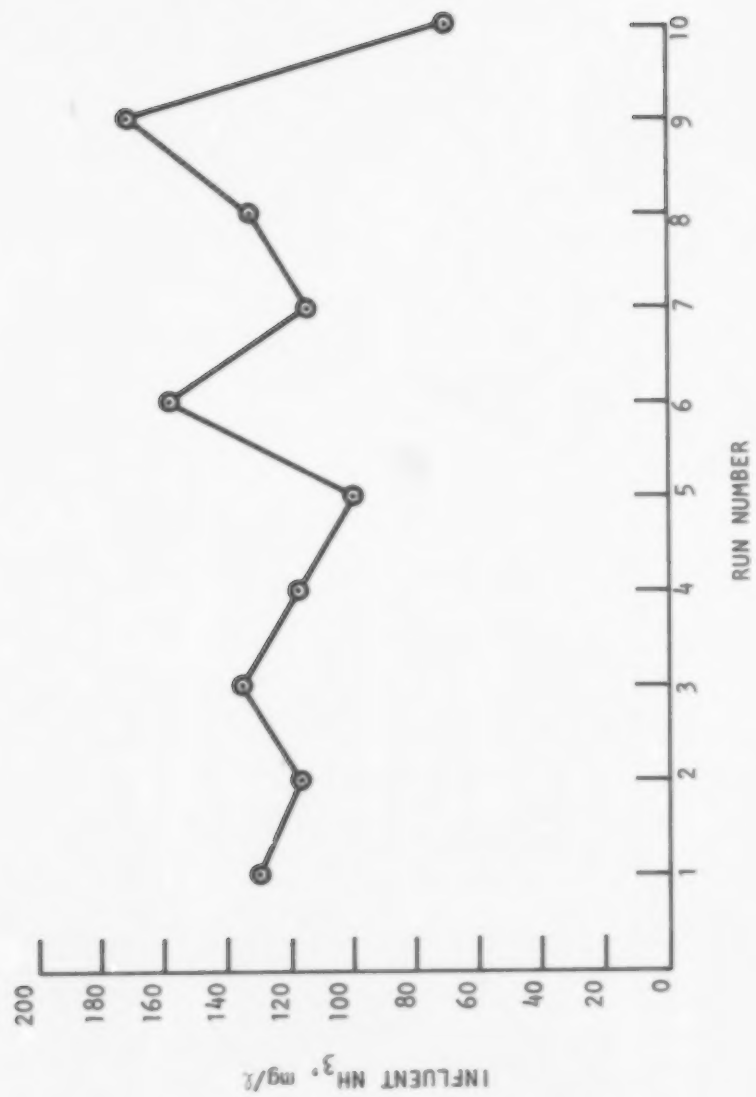


Figure 59. Reverse osmosis brine variation.

The BAT levels for the brine when discharged along with the RO product are different than the limits for treating raw wastewater because the limits are established in terms of kg/1,000 kg product (lb/1,000 lb). The RO product would make up 75 percent of the discharge stream while the brine would constitute 25 percent. The lower BAT parameter concentrations in the RO product allows higher concentrations in the brine without exceeding the limitations. Table 60 presents the calculated concentrations of the various BAT parameters that the ozonated brine must meet when discharged with the RO product. As can be seen, all BAT limitations, except for ammonia, are easily met. The study indicates that the ammonia limitation could be achieved.

TABLE 60. BAT LIMITS FOR DISCHARGE OF OZONATED BRINE AND RO PRODUCT

Parameter	Concentration, mg/l		
	RO Product ^a	Ozonated brine	Total
TSS	3	99	27
CN	0.01	1.054	0.271
Phenol	0.054	2.01	0.543
NH ₃ as N	9	16.40	10.85
S	0.2	0.736	0.334
F	0.40	85.64	21.71
pH	-	-	6.0-9.0

a. Projected water quality from pilot study.

Another option would be to reuse the reverse osmosis product and discharge the treated brine alone. The high quality RO product could be reused as makeup water for the blast furnace gas scrubber system or in a different part of the mill and reduce the quantity of lake water required. The low pollutant concentrations in the product water would not significantly increase the system blowdown pollutant concentrations or blowdown frequency. An average of 4,050 m³/day (1.07 mgd) of product water could be reused at the research site. The BAT limits for the treated brine would be significantly higher because of the lower volume discharges. The BAT limits for discharge of the ozonated brine only are presented on the following page.

The untreated reverse osmosis brine meets all limits except the ammonia limit. Five runs, numbers 1, 7, 8, 9, and 10, met the 43.3 mg/l ammonia discharge limit. Run 10 can be eliminated from consideration because the brine ammonia concentration was atypically low and data to determine the amount of ozone used could not be obtained. Table 61 presents the effluent quality and important operational parameters for the remaining four runs. The effluent pH outside the 6-9 range in Run 9 could easily be lowered to a value within the range. Better pH control or a lower set point on the pH

BAT LIMITS FOR DISCHARGE OF OZONATED BRINE ONLY

Parameter	Concentration, mg/l
TSS	108
CN _A	1.084
Phenol	2.172
NH ₃ as N	43.4
S	1.336
F	86.84
pH	6.0-9.0

TABLE 61. RO BRINE OZONATION RUNS MEETING BAT LIMITATIONS

	Date			
	3/7	3/13	3/14	3/15
Run number	1	7	8	9
Contact time, minutes	120	300	400	400
O ₃ applied, mg/l	991	1704	1877	3000
O ₃ utilized, mg/l	760	665	747	1465
Feed NH ₃ as N, mg/l	130.5	115	133	172
pH	6.90	6.30	9.90	8.50
SS, mg/l	38	78	53	65
Phenol, mg/l	<0.001	0.007	0.023	0.56
NH ₃ as N, mg/l	39	42	42	30 ^a
NO ₃ as N, mg/l	-	73	85	78
TP as P, mg/l	-	-	<0.02	-
CN _A , mg/l	<0.01	<0.01	<0.01	<0.01
CN, mg/l	<0.01	<0.01	<0.01	<0.01
F, mg/l	17.4	27.6	9.1	14.0
TOC, mg/l	22	4	6	4
Color, APHA units	0	-	0	0
Cond., µmhos/cm	14500	13500	13000	14500
TDS, mg/l	9795	10626	8353	9652
Cl, mg/l	4925	5600	4230	4830
O ₃ /NH ₃ , mg/mg	8.31	9.11	8.21	10.32

a. On-site measurement.

controller during neutralization would accomplish this. The high proportion of ozone utilized in Run 1 cannot be explained from the data. It is apparent that the higher transfer efficiency made meeting the BAT limit possible. This run confirms that shorter contact times can produce effluents of similar quality to longer contact times if the transfer efficiency is improved. The increase in TDS (close to 43 percent) during ozonation is due to the addition of dissolved solids during the pH adjustment

and neutralization. A large part of the increase is the chloride ions from the hydrochloric (HCl) acid used to neutralize the treated brine. The ratios of ozone required/ammonia removed are more consistent than the filtered wastewater study. The bench study predicted a ratio of 15.6 using a brine with lower ammonia but higher cyanide and phenol concentrations.

Mechanical and Operational Problems

Some mechanical problems occurred during the test period. A change in the type of feed pump used in the system had to be made because of the formation of large quantities of carbonate and hydroxide floc when the wastewater pH was raised to 11.5-12.0. This floc clogged the impeller of the pump provided with the unit. This situation was corrected by substituting a sump pump that was less susceptible to clogging than the original pump. A full-scale system would have to allow for the floc formation. At various times during the test program minor ozone leaks occurred in the ozone generator and gas piping systems. These leaks were detected and repaired in the field. Ozone leaks would not be a major problem in a full size system.

No major operational problems were encountered during the test program.

Full Size System Design

Filtered Wastewater Ozonation--

Cost estimates were prepared by an ozonation system manufacturer for three full-scale designs, based on the test program results. The cost estimates were prepared for system sizes of 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd), and 5,678 m³/day (1.5 mgd). The following cost guidelines were used:

1. Costs not to include pretreatment.
2. Manpower requirements not to include supervision or laboratory costs.

Costs were estimated for three configurations of ozonation systems, air feed, once through oxygen, and oxygen feed with recycle.

Each configuration contains the basic equipment, non-lubricated compressors with aftercoolers and moisture separators, heatless desiccant air dryers, ozonators, transformer reactors, 37.9 m³ (10,000 gal.) covered contact tank, ozone analyzer, ozone destruct units, and all necessary instrumentation and motor controls. The oxygen feed systems require liquid oxygen (LOX) storage tanks and vaporizers. The once through oxygen system would not require compressors, aftercoolers, or desiccant dryers. The number and capacity of the various components depends on the system size. A description of the operation of an air feed system will illustrate an ozonation system flow schematic. Ambient air is compressed through a non-lubricated compressor to 7.03 kg/cm² (100 psig). The compressed gas is cooled by watercooled aftercoolers and sent to compressed air receivers. The air is then sent through heatless desiccant dryers to lower the dew point to approximately

-51°C (-60°F). The 7.03 kg/cm² (100 psig) compressed air is regulated to the ozone generators at 1.05 kg/cm² (15 psig) by a pressure reducing valve. Ozone is produced to approximately 1 percent by weight in a high voltage electric discharge. The ozone stream is sent through ceramic diffusers at approximately 693 kg (1,529 lb) of ozone per day and a dosage of 733 mg/l of filtered wastewater. The ozone is in contact with the wastewater for approximately 60 minutes. The off-gas from the contactor is monitored by an ozone analyzer for control of ozone dosage. The off-gas from the contactor is sent through ozone destruction units and vented to the atmosphere.

Table 62 presents the cost estimates for air feed ozonation systems of 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd), and 5,678 m³/day (1.5 mgd) capacities. Costs for pH adjustment and neutralization including feed pumps, control equipment, storage tanks, and chemicals were added to the vendor estimates. A 30 day chemical storage capacity was assumed and the daily usage rates of NaOH and H₂SO₄ were based on the rates determined during the study.

Table 63 presents cost estimates for once through oxygen ozonation systems. Comparison with Table 62 illustrates that the cost per volume of wastewater treated is significantly lower for the air feed system although the oxygen feed system would be more economical for larger system capacities. Table 64 lists the cost estimates for oxygen feed with recycle systems. This system configuration would also be the most economical for large system capacities.

Space requirements vary for the system configurations. Table 65 compares area requirements of the three system configurations. As can be seen, space requirements are quite similar for the air and once through oxygen systems. The system using oxygen with recycle has a larger space requirement because it contains compressors and LOX storage tanks.

Reverse Osmosis Brine Ozonation--

Cost estimates were prepared by an ozonation system vendor for system capacities of 238 m³/day (0.063 mgd), 712 m³/day (0.188 mgd), and 1,419 m³/day (0.375 mgd). The vendor chose an air feed system for the 238 m³/day (0.063 mgd) system and oxygen feed with recycle for the 712 m³/day (0.188 mgd) and 1,419 m³/day (0.375 mgd) systems. The contactor tank for the air feed system provided a 14 hour contact time. The two oxygen feed systems utilized contactors providing 6.5 hr contact times. The higher concentration of ozone in the gas stream permits shorter contact time for oxygen feed systems. The applied ozone dosage for all three systems was set at 708 mg/l of wastewater based on the pilot scale study. The cost estimates for the reverse osmosis brine ozonation systems are given in Table 66. The significant difference in unit cost between the air feed and oxygen feed systems can be attributed to the economy of scale associated with larger system capacities.

The space requirements for the three systems are presented in Table 67.

TABLE 62. AIR FEED OZONATION SYSTEM COST

Design Flow Rate, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
<u>Capital Costs, \$^a</u>			
Delivered equipment total	\$ 877,000	\$ 2,375,000	\$ 4,533,000
Equipment installation	351,000	950,000	1,813,000
Installed Equipment Total	<u>\$1,228,000</u>	<u>\$ 3,325,000</u>	<u>\$ 6,346,000</u>
Process piping	369,000	998,000	1,904,000
Instrumentation	61,000	166,000	317,000
Capital Equipment Total	<u>\$1,658,000</u>	<u>\$ 4,489,000</u>	<u>\$ 8,567,000</u>
Facilities & Fees	3,069,000	8,304,000	15,848,000
Capital Cost Total	<u>\$4,727,000</u>	<u>\$12,793,000</u>	<u>\$24,415,000</u>
<u>Operating Costs, \$/Year</u>			
Manpower	\$ 54,000	\$ 54,000	\$ 54,000
Electricity ^b	223,000	670,000	1,340,000
Chemical ^c	45,000	136,000	271,000
Cooling water ^d	71,000	213,000	426,000
Maintenance ^e	83,000	224,000	428,000
Amortized capital ^f	441,000	1,196,000	2,282,000
Annual Cost, \$/Year	<u>\$ 917,000</u>	<u>\$ 2,493,000</u>	<u>\$ 4,801,000</u>
\$/3.785 m ³ (\$/1,000 gal.)	\$10.05	\$9.11	\$8.77

- a. See Appendix Tables A-4 and A-5 for basis of estimate.
b. \$0.03/KWH
c. pH adjustment NaOH, \$0.125/l (\$0.475/gal.), neutralization H₂SO₄, \$0.097/l (\$0.366/gal.)
d. \$0.00008/l (\$0.0003/gal.)
e. 5% per year of capital equipment cost.
f. 6.875% interest, 20 year equipment life.

TABLE 63. ONCE THROUGH OXYGEN OZONATION SYSTEM COSTS

Design Flow Rate, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Capital Costs^a			
Delivered equipment total	\$ 537,000	\$ 1,400,000	\$ 2,633,000
Equipment installation	215,000	560,000	1,053,000
Installed Equipment Total	\$ 752,000	\$ 1,960,000	\$ 3,686,000
Process piping	226,000	588,000	1,106,000
Instrumentation	38,000	98,000	184,000
Capital Equipment Total	\$1,016,000	\$ 2,646,000	\$ 4,976,000
Facilities & Fees	1,881,000	4,896,000	9,207,000
Capital Cost Total	\$2,897,000	\$ 7,542,000	\$14,183,000
Operating Costs, \$/Year			
Manpower	\$ 54,000	\$ 54,000	\$ 54,000
Electricity ^b	113,000	339,000	683,000
Chemical ^c	45,000	136,000	271,000
Oxygen ^d	446,000	1,340,000	2,680,000
Cooling water ^e	47,000	142,000	284,000
Maintenance ^f	51,000	132,000	249,000
Amortized capital ^g	271,000	705,000	1,326,000
Annual Cost, \$/Year	\$1,027,000	\$ 2,848,000	\$ 5,547,000
\$/3.785 m ³ (\$/1,000 gal.)	\$11.25	\$10.40	\$10.13

- a. See Appendix Tables A-4 and A-5 for basis of estimate.
b. \$0.03/KWH.
c. pH adjustment NaOH, \$0.125/l (\$0.475/gal.); neutralization H₂SO₄, \$0.097/l (\$0.366/gal.).
d. \$0.09/kg (\$0.04/lb).
e. \$0.00008/l (\$0.0003/gal.).
f. 5% per year of capital equipment cost.
g. 6.875% interest, 20 year equipment life.

TABLE 64. OXYGEN FEED WITH RECYCLE OZONATION SYSTEM COSTS

Design Flow Rate m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Capital Costs^a			
Delivered equipment total	\$ 767,000	\$ 2,075,000	\$ 3,933,000
Equipment installation	307,000	830,000	1,573,000
Installed Equipment Total	\$1,074,000	\$ 2,905,000	\$ 5,506,000
Process piping	322,000	872,000	1,652,000
Instrumentation	54,000	145,000	275,000
Capital Equipment Total	\$1,450,000	\$ 3,922,000	\$ 7,433,000
Facilities & Fees	2,683,000	7,255,000	13,752,000
Capital Cost Total	\$4,133,000	\$11,177,000	\$21,185,000
Operating Costs, \$/Year			
Manpower	\$ 54,000	\$ 54,000	\$ 54,000
Electricity ^b	142,000	433,000	854,000
Chemicals ^c	45,000	136,000	271,000
Oxygen ^d	252,000	755,000	1,510,000
Cooling water ^e	57,000	170,000	341,000
Maintenance ^f	72,000	196,000	372,000
Amortized capital ^g	386,000	1,045,000	1,980,000
Annual Cost \$/Year	\$1,008,000	\$ 2,789,000	\$ 5,382,000
\$/3,785 m ³ (\$/1,000 gal.)	\$11.05	\$10.19	\$9.83

- a. See Appendix Tables A-4 and A-5 for basis of estimate.
b. \$0.03/KWH.
c. pH adjustment NaOH, \$0.125/l (\$0.475/gal.); neutralization H₂SO₄, \$0.097/l (\$0.366 gal.).
d. \$0.09/kg (\$0.04/lb).
e. \$0.00008/l (\$0.0003/gal.).
f. 5% per year of capital equipment cost.
g. 6.875% interest, 20 year equipment life.

TABLE 65. OZONATION SYSTEM SPACE REQUIREMENTS

System configuration	Space requirements, m ² (ft ²)		
	Space capacity, m ³ /day (mgd)		
	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Air feed	34 (364)	102 (1,092)	204 (2,184)
Once through oxygen	36 (384)	108 (1,152)	216 (2,304)
Oxygen with recycle	51 (546)	153 (1,638)	306 (3,276)

TABLE 66. REVERSE OSMOSIS BRINE OZONATION SYSTEM COSTS

System Size, m ³ /day (mgd)	238 (0.063)	712 (0.188)	1,419 (0.375)
System Configuration	Air Feed	Oxygen With Recycle	Oxygen With Recycle
Capital Costs^a			
Delivered equipment total	\$ 529,000	\$ 680,000	\$1,314,000
Equipment installation	212,000	272,000	526,000
Installed equipment total	\$ 741,000	\$ 952,000	\$1,840,000
Process piping	222,000	286,000	552,000
Instrumentation	37,000	48,000	92,000
Capital Equipment Total	\$1,000,000	\$1,286,000	\$2,484,000
Facilities & Fees	1,850,000	2,380,000	4,595,000
Capital Cost Total	\$2,850,000	\$3,666,000	\$7,079,000
Operating Costs, \$/Year			
Manpower	\$ 18,000	\$ 18,000	\$ 18,000
Electricity ^b	39,000	58,000	118,000
Chemicals ^c	22,000	65,000	130,000
Oxygen ^d	-	183,000	366,000
Cooling water ^e	16,000	20,000	43,000
Maintenance ^f	50,000	64,000	124,000
Amortized capital ^g	266,000	343,000	662,000
Annual Cost, \$/Year	\$ 411,000	\$ 751,000	\$1,461,000
\$/3.785 m ³ (\$/1,000 gal.)	17.87	10.94	10.67

a. See Appendix Tables A-4 and A-5 for basis of estimate.

b. \$0.03/KWH.

c. pH adjustment NaOH, \$0.125/l (\$0.475/gal.); neutralization H₂SO₄, \$0.097/l (\$0.366/gal.).

d. \$0.09/kg (\$0.04/lb).

e. \$0.00008/l (\$0.0003/gal.).

f. 5% per year of equipment cost.

g. 6.875% interest, 20 year equipment life.

TABLE 67. BRINE OZONATION SYSTEM SPACE REQUIREMENTS

System capacity, m ³ /day (mgd)	Space requirements, m ² (ft ²)
238 (0.063)	41 (446)
712 (0.188)	62 (670)
1,419 (0.375)	94 (1,016)

RESULTS OF ACTIVATED CARBON DECHLORINATION

Introduction

Granular bituminous activated carbon systems were used to reduce residual chlorine following alkaline chlorination. Carbon dechlorination was investigated as part of two treatment trains. The first train consisted of alkaline chlorination of the raw wastewater, chemical treatment, clarification, and dechlorination. The second treatment train included chemical treatment and clarification of the raw wastewater, filtration, treatment by reverse osmosis (RO), alkaline chlorination of the RO brine, and dechlorination.

Alkaline Chlorinated Raw Wastewater

The first phase of the study utilized two carbon columns operating in an upflow, series mode, providing a contact time of 20 minutes. One carbon column, providing an 8.5 minute contact time, was used for the remainder of the study. The carbon columns were operated in an upflow configuration to minimize the problems associated with the evolution of carbon monoxide (CO) gas during dechlorination. These problems include carbon bed separation and gas bubble formation. Table 68 presents the data obtained during the study. During the first phase of the test program, the average influent chlorine concentration, expressed as total available residual chlorine, was 49.8 mg/l. The 20 minute contact time produced an effluent chlorine concentration of 0.88 mg/l or a removal efficiency of 98.2 percent. Residual chlorine measurements were performed in the field on grab samples taken throughout the carbon bed. The average influent chlorine concentration during the portion of the study when one carbon column was in use was 74.6 mg/l. The 8.5 minute contact time produced an effluent chlorine concentration of 4.27 mg/l, or a removal efficiency of 94.3 percent.

Currently, no BAT effluent guidelines have been proposed for residual chlorine in treated blast furnace scrubber wastewater. Municipal effluent guidelines call for less than 1 mg/l total available chlorine. Of this, only 0.2-0.3 mg/l may be free chlorine (HOCl and OCl^-). If similar guidelines were established for blast furnace scrubber wastewater, activated carbon dechlorination could meet these guidelines. The pilot scale investigation indicated that a 20 minute contact time would be required.

For this particular application, an activated carbon dechlorination system would also be required to reduce the concentration of phenol in the wastewater. The BAT phenol effluent guideline is 0.543 mg/l at this research site. The influent phenol concentration to the dechlorination system never exceeded 0.5 mg/l during the test program. However, the phenol concentration in the blast furnace blowdown exceeded the limitation on several occasions while the mobile treatment units were at the research site. For this reason, a full-scale activated carbon dechlorination system would also be required to have the capability for phenol adsorption.

The BAT suspended solids effluent guideline for this specific steel mill

TABLE 68. RAW WASTEWATER DECHLORINATION

Date	Contact time, minutes	Influent total available chlorine, mg/l	Effluent total available chlorine, mg/l
2/5	20	21.8	1.77
2/6	20	21.3	0.88
2/9	20	106.3	0.0
2/12	8.5	97.4	1.77
2/13	8.5	25.0	1.90
2/13	8.5	39.0	1.30
2/14	8.5	40.5	1.80
2/14	8.5	70.4	3.50
2/15	8.5	79.3	1.77
2/19	8.5	113	7.30
2/20	8.5	113	7.10
2/21	8.5	99.2	8.80
2/22	8.5	69	7.50

is 27 mg/l. The average influent suspended solids during the study was 26 mg/l and the effluent concentration was 17 mg/l. A carbon dechlorination system utilizing an upflow mode of operation is not designed for suspended solids removal. If the chemical treatment and clarification system prior to the dechlorination system cannot effectively reduce the suspended solids concentration to below BAT limits, an additional process such as pressure filtration may be required preceding carbon dechlorination.

Alkaline Chlorinated Reverse Osmosis Brine

The alkaline chlorinated reverse osmosis brine dechlorination study used one carbon column operated in an upflow configuration. Contact times ranged from 19.5-44.9 minutes. The high dissolved solids of the RO brine required that a modified Iodometric Method I [Standard Methods (2)] be used to prevent interference by ferric, manganese, and nitrite ions. The modified method did not include acidification with acetic acid. Table 69 presents the data obtained during the study. The average total available chlorine concentration in the influent was 28.1 mg/l. The influent free chlorine averaged 16.3 mg/l. The effluent total available and free chlorine concentrations averaged less than 0.1 mg/l. This corresponds to a total available

TABLE 69. REVERSE OSMOSIS BRINE DECHLORINATION

Date	Contact time, minutes	Influent chlorine, mg/l		Effluent chlorine, mg/l	
		Total available	Free	Total available	Free
3/25	21.4	45	10	<0.1	<0.1
3/25	19.5	-	25	<0.1	<0.1
3/25	44.9	39.8	15	<0.1	0.3
3/26	37.4	<0.1	50	<0.1	<0.1
3/26	24.9	53	2.5	<0.1	<0.1
3/26	24.9	26.5	10	<0.1	0.1
3/27	40.8	4.4	1.3	<0.1	<0.1

chlorine removal of greater than 99.6 percent and a free chlorine removal of greater than 99.4 percent. These high removals permitted the municipal standards for total available and free chlorine to be met.

The dechlorination system influent suspended solids concentration averaged 28 mg/l. The BAT suspended solids effluent guideline for discharge of the alkaline chlorinated RO brine is 108 mg/l because of the smaller volume of wastewater.

Full-scale System Design

The data generated in the pilot study at the blast furnace research site was used for system scale-up to full size installations. Full-scale system costs were based on information submitted by a bituminous carbon vendor. Several assumptions were made in system scale-up and costing. These assumptions are listed in Table 70.

Raw Wastewater Dechlorination--

The full-scale raw wastewater dechlorination systems description and operation characteristics are presented in Table 71. All proposed systems would utilize two lined carbon steel adsorbers. One adsorber would be on stream with the second adsorber held in reserve. All dechlorination systems would be arranged in a fixed bed, upflow operating mode. The carbon exhaustion rate for all proposed systems would be 0.24 kg/1,000 liters (2.0 lb/1,000 gal.) of wastewater treated. The exhausted carbon would be replaced with virgin carbon when the effluent chlorine residual exceeded the maximum concentration. At that time, the adsorber kept in reserve would be put on stream and the spent carbon in the second adsorber would be replaced with virgin carbon. Spent carbon could not be regenerated and would require disposal.

TABLE 70. ASSUMPTIONS USED TO DEVELOP COST AND DESIGN FIGURES FOR THE CARBON DECHLORINATION SYSTEM

1. Electricity costs: \$0.03/KWH		
2. Design feed flow:		
	Raw Wastewater	Reverse Osmosis Brine
	946 m ³ /day (0.25 mgd)	237 m ³ /day (0.0625 mgd)
	2,839 m ³ /day (0.75 mgd)	710 m ³ /day (0.1875 mgd)
	5,678 m ³ /day (1.5 mgd)	1,419 m ³ /day (0.375 mgd)
3. Capital amortization: 6.875%		
	20 year service life	
4. \$18,000/year skilled labor		
\$16,000/year unskilled labor		
5. The costs will not include wastewater storage facilities or distribution systems.		
6. Manpower requirements will not include supervision or laboratory costs.		
7. Carbon saturation at 0.209 gm Cl ₂ /gm of carbon.		
8. Contact time of 20 minutes.		

Table 72 presents the cost estimates for the proposed systems. The capital costs of the activated carbon systems include the adsorption system and a virgin carbon storage system. Installation costs are included in the capital costs. The capital cost of the adsorption systems include pumps, adsorbers, flow control, piping, electrical work, foundations, winterization, the initial carbon fill, and indirect costs. The virgin carbon storage systems are designed to hold a 30 day supply. The capital costs include all tanks, blow cases, pumps, piping, electrical work, foundations, winterization, the initial carbon fill, and indirects.

Reverse Osmosis Brine Dechlorination--

The full-scale RO brine dechlorination systems description and operating characteristics are given in Table 73. All proposed systems would contain two lined carbon steel adsorbers. One adsorber would be on stream with the second adsorber held in reserve. All adsorbers would be operated in a fixed bed, upflow mode. The expected carbon exhaustion rate is 0.36 kg/1,000 liters (2.99 lb/1,000 gal.) of wastewater dechlorinated.

TABLE 71. RAW WASTEWATER DECHLORINATION
SYSTEM DESCRIPTION AND OPERATING CHARACTERISTICS

System size, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Feed chlorine, (mg/l total available Cl ₂)	50	50	60
Effluent chlorine, (mg/l total available Cl ₂)	1.0	1.0	1.0
System operation	Upflow	Upflow	Upflow
Number of adsorbers	2	2	2
Number of adsorbers on stream	1	1	1
Adsorber dimension, m (ft)	2.4x4.3 (8x14)	3.7x5.8 (12x19)	4.9x6.4 (16x21)
Carbon depth, m (ft)	2.81 (9.23)	3.78 (12.4)	4.24 (13.9)
Contact time, minutes	20	20	20
Carbon type		bituminous	
Carbon density, dry, kg/m ³ (lb/ft ³)	500 (31.2)	500 (31.2)	500 (31.2)
Carbon density, backwashed and drained, kg/m ³ (lb/ft ³)	424 (26.5)	424 (26.5)	424 (26.5)
Volume of carbon in initial fill, m ³ (ft ³)	26.3 (928)	79.3 (2,800)	159 (5,600)
Weight of carbon in initial fill, kg (lb)	13,200 (29,000)	39,700 (87,400)	79,400 (174,700)
Impurity loading, gm/gm	0.209	0.209	0.209
Carbon dosage, kg/1,000 (lb/1,000 gal.)	0.24 (2.0)	0.24 (2.0)	0.24 (2.0)
Exhaustion rate, kg/day (lb/day)	227 (500)	680 (1,500)	1,360 (3,000)
Initial carbon fill price, \$/kg (\$/lb)	1.56 (0.71)	1.56 (0.71)	1.56 (0.71)
Make-up carbon fill price, \$/kg (\$/lb)	1.56 (0.71)	1.56 (0.71)	1.56 (0.71)

TABLE 72. RAW WASTEWATER DECHLORINATION SYSTEM
COST ESTIMATES

System size, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Capital Cost^a			
Adsorption system	\$ 253,000	\$ 430,000	\$ 702,000
Carbon storage system ^b	71,000	144,000	225,000
Capital Equipment Total	\$ 324,000	\$ 574,000	\$ 927,000
Facilities & Fees	696,000	1,233,000	1,994,000
Total Plant Cost	\$1,020,000	\$1,807,000	\$2,921,000
Operating Costs, \$/Year			
Operator ^c	\$ 12,000	\$ 16,000	\$ 16,000
Carbon make-up ^d	130,000	389,000	777,000
Maintenance ^e	16,000	29,000	46,000
Amortization of capital ^f	95,000	169,000	273,000
Total Annual Costs	\$ 253,000	\$ 603,000	\$1,112,000
\$/3,785 & (\$/1,000 gal.)	2.77	2.20	2.03

- a. See Appendix Tables A-4 and A-5 for basis of estimate.
- b. Storage for 30 day supply.
- c. Unskilled at \$16,000/year rate.
- d. \$1.56 kg (\$0.61/lb).
- e. 5% per year of capital equipment cost.
- f. 20 year service life, 6.875% interest.

TABLE 73. REVERSE OSMOSIS BRINE DECHLORINATION SYSTEM
DESCRIPTION AND OPERATING CHARACTERISTICS

System size, m ³ /day (mgd)	237 (0.0625)	710 (0.1875)	1,419 (0.375)
Feed chlorine, (mg/l total available Cl ₂)	75	75	75
Effluent chlorine, (mg/l total available Cl ₂)	1.0	1.0	1.0
System operation		upflow	
Number of adsorbers	2	2	2
Number of adsorbers on stream	1	1	1
Adsorber dimension, m (ft)	1.5x2.7 (5x9)	2.4x3.3 (8x11)	3.0x4.3 (10x14)
Carbon depth, m (ft)	1.80 (5.91)	2.11 (6.92)	2.70 (8.86)
Contact time, minutes	20	20	20
Carbon type		bituminous	
Carbon density, dry, kg/m ³ (lb/ft ³)	500 (31.2)	500 (31.2)	500 (31.2)
Carbon density, backwashed and drained, kg/m ³ (lb/ft ³)	424 (26.5)	424 (26.5)	424 (26.5)
Volume of carbon in initial fill, m ³ (ft ³)	6.57 (232)	19.7 (696)	39.4 (1,392)
Weight of carbon in initial fill, kg (lb)	3,300 (7,200)	9,900 (21,700)	19,700 (43,400)
Impurity loading, gm/gm	0.209	0.209	0.209
Carbon dosage, kg/1,000 l (lb/1,000 gal.)	0.36 (2.99)	0.36 (2.99)	0.36 (2.99)
Exhaustion rate, kg/day (lb/day)	84.8 (187)	254 (561)	509 (1,120)
Initial carbon fill price, \$/kg (\$/lb)	1.56 (0.71)	1.56 (0.71)	1.56 (0.71)
Make-up carbon fill price, \$/kg (\$/lb)	1.56 (0.71)	1.56 (0.71)	1.56 (0.71)

The cost estimates for the proposed systems are shown in Table 74. The system capital costs include the adsorption system and a virgin carbon storage system.

The adsorption systems capital costs include installation, pumps, adsorbers, flow control, piping, electrical work, foundations, winterization, the initial carbon fill, and indirect costs. The virgin carbon storage systems are designed to hold a 30 day supply. The capital costs include all tanks, blow cases, pumps, piping, electrical work, foundations, winterization, the initial carbon fill, and indirects.

System Space Requirements

A summary of the activated carbon dechlorination system space requirements is presented in Table 75. The space requirements were developed from vendor information for both raw wastewater and reverse osmosis brine dechlorination.

TABLE 74. REVERSE OSMOSIS BRINE DECHLORINATION
SYSTEM COST ESTIMATES

System size, m ³ /day (mgd)	237 (0.0625)	710 (0.1875)	1,419 (0.375)
Capital Cost^a			
Adsorption system	\$207,000	\$223,000	\$ 299,000
Carbon storage system ^b	49,000	77,000	120,000
Capital Equipment Total	\$256,000	\$300,000	\$ 419,000
Facilities & Fees	551,000	645,000	901,000
Total Plant Cost	\$807,000	\$945,000	\$1,320,000
Operating Costs, \$/Year			
Operator ^c	\$ 12,000	\$ 12,000	\$ 12,000
Carbon make-up ^d	48,000	145,000	290,000
Maintenance ^e	40,000	47,000	66,000
Amortization of capital ^f	75,000	88,000	123,000
Total Annual Costs	\$175,000	\$292,000	\$ 491,000
\$/3,785 _l (\$/1,000 gal.)	7.67	4.27	3.59

- a. See Appendix Tables A-4 and A-5 for basis of estimate.
b. Storage for 30 day supply.
c. Unskilled at \$16,000/year rate.
d. \$1.56/kg (\$0.71/lb).
e. 5% per year of capital equipment cost.
f. 20 year service life, 6.875% interest.

TABLE 75. ACTIVATED CARBON DECHLORINATION
SYSTEM SPACE REQUIREMENTS

Wastewater	System size, m ³ /day (mgd)	Space requirements, m ² (ft ²)
Raw wastewater	946 (0.25)	255 (2,750)
	2,839 (0.75)	297 (3,200)
	5,678 (1.5)	372 (4,000)
Reverse osmosis brine	237 (0.0625)	223 (2,400)
	710 (0.1875)	241 (2,600)
	1,419 (0.375)	265 (2,850)

SECTION 8

PRIORITY POLLUTANT DISCUSSION

INTRODUCTION

Priority pollutant removal from the blast furnace blowdown was evaluated for the ozonation, alkaline chlorination and reverse osmosis (RO) processes. Detailed descriptions of the research site (Section 4) and the treatment technologies (Section 6) have been provided.

The priority pollutants arose from a June 7, 1978 court settlement involving the EPA and several environmental plaintiffs who were concerned that the EPA was not fulfilling the mandate outlined in the Federal Water Pollution Control Act P.L. 92-500. The result of the suit, commonly known as the "EPA Consent Decree", was the creation of a list of toxic and/or hazardous materials for which technology based effluent limitations would be required. The original Consent Decree required that source performance standards and pretreatment standards for 65 pollutants be developed for 21 industrial categories, including iron and steel manufacturing. The list of pollutants was later expanded to 129 materials which are divided into eight categories; metals, asbestos, total cyanides, pesticides, acid extractable organics, base/neutral extractable organics, volatile organics and total phenols. The EPA then implemented a three phase program to identify priority pollutants in industrial effluents. Phase I was intended as a screening process, Phase II verification and Phase III monitoring. All sampling and analyses for priority pollutants in this project were done under protocol developed for either Phase I or Phase II of the EPA program. The sampling was performed by Rexnord Environmental Research Center personnel and the analyses were done by TRW Defense and Space Systems Group. A draft copy of the TRW report containing all sampling and analytical information is provided in Appendix I.

The evaluated treatment technologies were all able to reduce total metal concentrations to less than 10 $\mu\text{g/l}$ for all the priority pollutant metals except zinc and selenium. The only organic priority pollutants reported were bis (2 ethylhexyl) phthalate (BEP) and di-n-butyl phthalate (DBP). DBP concentrations were below the verification limit (10 $\mu\text{g/l}$). The precision and accuracy for BEP did not allow a quantitative evaluation of treatment technology effectiveness. All treatment technology process trains removed cyanide to <20 $\mu\text{g/l}$. Some discrepancies exist between the colorimetric method for total phenols and the GC/MS acid extractable fraction, which are phenolic based compounds.

PRESENTATION OF RESULTS

Ozonation

The results of the priority pollutant analysis for the ozonation process are given in Table 76. Priority pollutant sampling was done on 20 January, 1979 during run number 14. The contact time was 171 minutes and the applied ozone dose was 1,074 mg/l of blast furnace wastewater. The final design values for the technology were 733 mg O_3 /l and 60 minutes contact time. The treatment process met BAT discharge limits for all NPDES monitored parameters.

All priority pollutant metals except zinc were reduced to $<10 \mu\text{g/l}$. Zinc was found in high concentrations (1.0 mg/l) in the carriage water as well as the blast furnace gas cleaning water. Carriage water at this site was Lake Michigan surface water which is used as makeup water in the recirculation system. Makeup water is added on an "as-needed" basis. The analysis program was not extensive enough to determine how much of the zinc was process generated and how much was due to makeup water.

Cyanide and phenols were removed to well below BAT limits. The only organic above verification level was BEP. Reported influent samples were 224 to 288 $\mu\text{g/l}$ (replicates of a single sample). Ozone effluent concentrations for a single sample were 562; 435; 395 $\mu\text{g/l}$, an apparent increase after ozonation. Carriage water values were from 80 to 115 $\mu\text{g/l}$. TRW reported spiked recoveries from 67 percent to 120 percent. These data are not sufficient to evaluate ozonation as a process for removing BEP.

Alkaline Chlorination

Priority pollutant samples for the alkaline chlorination technology were taken on 15 February, 1979 during run number 13. The data are summarized in Table 77. The chlorine to ammonia ratio during the run was very near that chosen for design considerations (9.6:1 vs. 10:1 for design). The high suspended solids value reported for the influent was the result of a temporary shutdown of the blast furnace scrubber water thickeners. Arsenic, lead and total organic carbon were also higher than the influents taken during the other sampling events. The other parameters were generally within the same concentration range for all three technologies. BAT limits were met for all parameters except suspended solids.

The alkaline chlorination process reduced all metals concentrations to $<10 \mu\text{g/l}$ except selenium and zinc. No explanation is available for the apparent fourfold increase in selenium through the process. Zinc was reduced 45 percent to 470 $\mu\text{g/l}$. Cyanide and phenol were both lower than 10 $\mu\text{g/l}$. BEP recoveries (surrogate recoveries) were 37 percent to 51 percent. The mean of three replicates for the influent sample was 58 $\mu\text{g/l}$. The effluent was 15 $\mu\text{g/l}$, indicating an apparent reduction through the process. However, these data allow only the most qualitative review of process efficiency.

TABLE 76. RESULTS OF PRIORITY POLLUTANT ANALYSES - OZONE PROCESS

<u>Operational parameters</u>									
Run #	Contact time, min.	O ₃ Dose, mg/l		O ₃ Gas temp.		O ₃ Gas press.		O ₃ Gas conc.	
		applied	utilized	OC	psi			mg/l	
14	171	1,074	480	37.2	15				13.95
<u>General analyses</u>									
	pH units	SS, mg/l	VSS, mg/l	NH ₃ -N, mg/l	CN, mg/l	CN _A , mg/l	S, mg/l	F, mg/l	Phenol, mg/l
Influent	7.7	160	17	29.9	0.33	0.33	N/A	12.5	0.025
Effluent	6.9	15	5	3.4	<0.01	<0.01	<0.2	12	0.005
BAT limits	N/A	24	N/A	9.6	3.0	N/A	0.29	19	0.48
<u>Priority pollutants</u>									
	As	Cr	Cu	Pb	Se	Zn	Hg	Cyanide, mg/l	
Raw	163	<10	32	189	16	580	0.4	0.208:0.278 (b)	
Effluent	<10	(a.)	14:<10 (b)	<10	<10	800	(a.)	0.002	
Carriage water	<10	(a.)	<10	<10	10	1,000	(a.)	0.005	

a. below verification limit

b. duplicate results

c. not applicable

TABLE 77. RESULTS OF PRIORITY POLLUTANT ANALYSES -
REVERSE OSMOSIS PROCESS

General analyses									
	pH	SS, mg/l	VSS, mg/l	NH ₃ -N, mg/l	Phenol, mg/l	CNA, mg/l	S, mg/l	F, mg/l	TOC, mg/l
Influent (a)	7.8	61	N/A ^c	43	0.033	N/A	N/A	9.8	7
Product	6.0	6	2	28	0.023	<0.01	N/A	0.65	4
Brine	7.5	21	N/A	133	0.228	0.95	N/A	25.1	24
BAT limits	N/A	32	N/A	12.9	0.64	0.32	N/A	26	N/A

Priority Pollutants									
	Metals, µg/l								Cyanide, mg/l
	As	Cr	Cu	Pb	Se	Zn	Hg		
Influent	179	21	24	161	<5	196	(b.)		0.298
Product	<10	6	2	(b.)	<5	<50	(b.)		0.015
Brine	11	4	1	2	<5	60	(b.)		1.210
Carriage water	<10	2	4	(b.)	<5	<50	(b.)		0.002

- a. Treatment train for R0 included chemical flocculation, precipitation, clarification and dual media filtration. Some pollutants were removed by these upstream processes. A mass balance of the R0 system from the presented data is not possible.
- b. Below verification limits.
- c. Not applicable.

Reverse Osmosis

The reverse osmosis system was operated at 75 percent recovery during priority pollutant sampling. Upstream clarification and filtration are primarily responsible for suspended solids removal from the influent (Table 78). The RO product water quality met all BAT requirements except ammonia. The brine failed to meet BAT limits for ammonia and cyanide and would require further treatment before discharge.

All reported metals were either less than 10 $\mu\text{g/l}$ or, in the case of zinc, less than 50 $\mu\text{g/l}$. Recoveries of the organic priority pollutant BEP were 46 to 69 percent. The concentrations of BEP were 13 $\mu\text{g/l}$ in the feed, 24 $\mu\text{g/l}$ in the brine and 89 $\mu\text{g/l}$ in the product. There is no explanation of why BEP should be higher in the product than in the brine or feed. The RO process affected a 95 percent removal of cyanide from the blast furnace blowdown waste. The RO brine concentration of 1.2 mg CN/l would have to be reduced by one half before meeting BAT limits. Seventy-eight percent of the cyanide in the brine was amenable to destruction by chlorination. The total phenol concentration in the brine was reported as 228 $\mu\text{g/l}$ by the colorimetric method. There was no confirmation of phenolic compounds by GC/MS analysis.

CONCLUSIONS

The priority pollutant metals, with the exception of zinc and selenium were removed to less than 10 $\mu\text{g/l}$ by all process technologies evaluated. Selenium was reported at 29 $\mu\text{g/l}$ in the alkaline chlorination process effluent only. For the other two process effluents it was 10 $\mu\text{g/l}$ or less. Zinc was found in high concentrations in the blast furnace recycle blowdown as well as the makeup water (carriage water). Only the reverse osmosis process was capable of removing zinc to less than 50 $\mu\text{g/l}$. It is unclear if the source of zinc is the blast furnace process or makeup water.

The only organic priority pollutant reported above verification levels was bis-(2 ethylhexyl) phthalate (BEP). Priority pollutants were sampled at this site in August of 1978. A summary of the three day sampling program is given in Table 79. BEP was reported on two of the three days monitored. The concentration on day No. 1 (the highest concentration) was three times higher than the highest BEP concentration found during this study and five times higher than the highest untreated blowdown wastewater values. BEP was reported in 91 percent of the industrial categories and 42 percent of the samples taken during the screening phase of the EPA's priority pollutant project. The chemical is either an artifact of the prescribed sampling and analytical methods or is a widely distributed organic pollutant. The highly variable nature of the state-of-the-art of BEP analysis does not allow for an analysis of its removal through the evaluated technologies. Taken literally, the data indicate BEP is increased by ozonation and RO and reduced by alkaline chlorination. No recommendation of treatment processes for BEP removal is possible at this time. The other reported organics present in August 1978 were not found in concentrations above 10 $\mu\text{g/l}$ during

TABLE 78. RESULTS OF PRIORITY POLLUTANT ANALYSES-
ALKALINE CHLORINATION PROCESS

Operational parameters		Contact time, min.		Chlorine:Ammonia		Chlorine residual, mg/l	
Run #	CN tank	NH ₃ tank	Clarifier	ratio	CN tank	NH ₃ tank	Clarifier
13	12	15	144	9.6:1	145	73	79

General analyses		pH	SS, mg/l	VSS, mg/l	NH ₃ -N, mg/l	Phenol, mg/l	CN _A , mg/l	S, mg/l	F, mg/l	TOC, mg/l
Influent		7.9	1,810	315	28.6	0.077	0.10	<0.2	14.6	280
AC Eff. (a.)		7.0 ^d	29	10	5.7	0.009	<0.01	<0.2	13.9	14
BAT limits		N/A	23	N/A	9.1	0.46	0.23	0.28	18.2	N/A

Priority pollutants		Metals, µg/l						Cyanide, mg/l
		As	Cr	Cu	Pb	Se	Zn	Hg
Influent		277	(b.)	16	1,140	7	860	<0.5
Chlorinator eff.		<10	(b.)	43	14	24	320	<0.5
AC eff. (a.)		<10	(b.)	<10; 10 (c.)	10	29	470	<0.5

- a. activated carbon effluent.
b. below verification limits.
c. duplicate analyses.
d. not applicable.

this study.

Cyanide was removed to well below BAT levels by all evaluated technologies.

TABLE 79. SUMMARY OF PRIORITY POLLUTANTS IN RECYCLE BLOWDOWN
REPORTED AUGUST, 1978

Parameter, $\mu\text{g/l}$

Day	Chloroform	BEP ^a	BBP ^b	DBP ^c	Anthracene	Phenathrene
1	N/A ^d	1,560	22	17	N/A	N/A
2	43	N/A	N/A	N/A	≤ 12	≤ 12
3	N/A	26	N/A	N/A	N/A	N/A

a. bis- (2 ethylhexyl) phthalate

b. butyl benzyl phthalate

c. di-n-butyl phthalate

d. not applicable

SECTION 9
SUMMARY AND EVALUATION

INTRODUCTION

This report portion summarizes the data which have previously been presented. For this portion of the report, the individual processes have been grouped into treatment systems capable of meeting the 1974 Best Available Technology (BAT) Economically Achievable levels for blast furnace scrubber wastewater.

During the study, blast furnace production averaged 10,919 metric tons/day (12,029 tons/day), with a scrubber water blowdown flow of 5,397 m³/day (1.426 mgd). Based on this information and the 1974 BAT guidelines, the allowable pollutant concentrations in the effluent from the plant investigated would be:

Parameter	BAT level	
	Blowdown	Reverse osmosis brine ^a
pH	6.0-9.0	6.0-9.0
Suspended solids, mg/l	27	108
Ammonia, mg/l	10.85	43.4
Cyanide-A, mg/l	0.271	1.08
Phenol, mg/l	0.543	2.17
Sulfide, mg/l	0.334	1.34
Fluoride, mg/l	21.71	86.8

a. Assuming reuse of the reverse osmosis product water.

The pilot studies were structured toward meeting the above guidelines. The results of the pilot program have indicated that alkaline chlorination, ozonation and reverse osmosis were all capable of producing an effluent acceptable for discharge which met the above BAT limitations. The discussions presented below will evaluate each of these systems from a standpoint of effluent quality, costs and space requirements.

The capital cost^a information presented in this report has been prepared from a minimum of flow sheet and equipment data. Using the definition found in Perry's "Chemical Engineers' Handbook" (20), this type of estimate would be termed a "study estimate", having an accuracy of ± 30 percent.

ALKALINE CHLORINATION TREATMENT TRAIN

The alkaline chlorination treatment train was successful in reducing influent concentrations of all parameters of concern to below BAT levels. The treatment train consisted of elevating the pH of the incoming wastewater to 11.0-11.5 with addition of sodium hypochlorite for oxidation of cyanide. The waste was then neutralized for ammonia removal. The exiting wastewater was dosed with chemical coagulants prior to clarification and the settled effluent was dechlorinated on activated carbon.

Performance Data

The alkaline chlorination treatment train was effective in reducing influent contaminant strengths to below BAT levels provided the $\text{Cl}_2:\text{NH}_3$ ratio was high enough. For design purposes, a ratio of $\text{Cl}_2:\text{NH}_3$ of 10:1 was selected. Performance data for a pilot run using this ratio are presented below in Table 80.

TABLE 80. ALKALINE CHLORINATION TREATMENT TRAIN
PERFORMANCE DATA

Run no.	Process	pH	SS mg/l	NH_3 mg/l	CN-A mg/l	Phenol mg/l	S^{2-} mg/l	F^- mg/l	Cl_2 mg/l
6	Influent	7.8	73	32.9	0.19	0.024	<0.2	13.5	0
	Alk. clar. effluent	7.1	137	25.9	0.10	-	-	-	42
	Clarifier effluent	7.3	16	0.48	<0.01	0.014	<0.2	12.5	21.3
	Act. car. effluent	6.8	6	1.0	<0.01	0.216	<0.2	12.5	1.3

a. All cost information is effective as of May 1979.

Comparing the above table and previously presented results to BAT requirements, it is obvious that this treatment train was very effective in treating the blast furnace scrubber blowdown. Although only one performance run is presented in Table 80 to eliminate repetition, it should be noted that several of the experimental runs met the BAT limits.

Cost Data

A summary of the costs associated with alkaline chlorination of the blast furnace scrubber blowdown stream is listed in Table 81. The systems shown have been designed to treat the wastewater to BAT levels. Designs are presented for flows of 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd) and 5,678 m³/day (1.5 mgd).

TABLE 81. ALKALINE CHLORINATION TREATMENT TRAIN COST SUMMARY

Size, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
<u>Capital Costs (\$)</u>			
Alkaline chlorination	\$ 421,000	\$ 533,000	\$ 611,000
Clarification	201,000	341,000	433,000
Total w/o Activated Carbon	\$ 622,000	\$ 874,000	\$1,044,000
Activated Carbon	1,020,000	1,807,000	2,921,000
Total w/Activated Carbon	\$1,642,000	\$2,681,000	\$3,965,000
<u>Operating Costs (\$/Yr)</u>			
Alkaline chlorination	\$ 273,000	\$ 602,000	\$1,009,000
Clarification	40,000	75,000	111,000
Total w/o Activated Carbon	\$ 313,000	\$ 677,000	\$1,120,000
Activated Carbon	253,000	603,000	1,112,000
Total w/Activated Carbon	\$ 566,000	\$1,280,000	\$2,232,000
<u>Operating Costs, \$/3,785 l (\$/1,000 gal.)</u>			
Alkaline chlorination	2.99	2.20	1.84
Clarification	0.44	0.27	0.20
Total w/o Activated Carbon	3.43	2.47	2.04
Activated Carbon	2.77	2.20	2.03
Total w/Activated Carbon	6.20	4.67	4.07

The alkaline chlorination system design assumes the use of chlorine gas and a 20 minute detention time for cyanide removal followed by a one minute ammonia removal step. A thirty day chemical stock tank storage has also been assumed. The clarification system assumes the use of ferric chloride and polymer. The design overflow rate is 22 m³/m²/day (530 gpm/ft²). The activated carbon unit is designed for a 20 minute wastewater detention time and will reduce the free chlorine concentration from 50 mg/l down to

one mg/l. It should be noted however, that unless dechlorination is required, activated carbon is not an essential part of the alkaline chlorination treatment system.

Space Requirements

Table 82 lists the space requirements for various size alkaline chlorination treatment systems. The space requirements presented include an allowance for the area between structures, space for interconnecting piping, etc. The information presented is preliminary and should be used mainly for comparative purposes. The actual space requirements will be determined by the final design criteria.

TABLE 82. ALKALINE CHLORINATION
TREATMENT TRAIN SPACE REQUIREMENTS

System capacity, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Space requirements m ² (ft ²)			
Alkaline chlorination	163 (1,750)	204 (2,200)	279 (3,000)
Clarification	51 (550)	130 (1,400)	288 (3,100)
Activated carbon	255 (2,750)	297 (3,200)	372 (4,000)
Total	469 (5,050)	631 (6,800)	938 (10,100)

OZONATION TREATMENT TRAIN

Ozonation was the second successful treatment technology investigated to reduce the blast furnace scrubber blowdown to below future BAT levels. The treatment train consisted of filtration, pH elevation to 10.5-11.5, ozonation, neutralization and clarification.

Performance Data

During the study, both a dual media filter and a magnetic filter were investigated for suspended solids and turbidity removal prior to ozonation. Dual media filtration removed 83 percent of the influent suspended solids and 75 percent of the feed turbidity. This compared to 70 percent SS removal and 32 percent turbidity reduction on the same feed water for the magnetic filter. Based on these performance parameters, the dual media filter provided a higher quality feedwater to the ozone process element

Preliminary ozonation studies indicated that the ammonia limitation was the critical factor in meeting the required discharge limitations for the oxidizable contaminants. Therefore, the ozone test study was structured toward meeting the ammonia limitation.

For the scrubber blowdown discharge, the prefiltered wastewater pH was elevated to 10.5-11.5 prior to entering the ozone contact columns. The total applied ozone dosage was varied from 0-1,500 mg/l. Contact times were adjusted between 60-240 minutes to determine optimum contactor retention times. Selected results of testing are shown in Table 83. The data indicated that ozone was effective in meeting BAT limits. One test run was conducted at elevated pH using compressed air. Results verified that ozonation rather than stripping was the mechanism for removing ammonia.

Suspended solids present in the neutralized ozonated effluent were easily removed by clarification. For the ozonated scrubber blowdown, polymer was added at a rate of 0.5 mg/l to improve settling. Suspended solids removal consistently averaged 95 percent or better at all flow rates investigated.

TABLE 83. BLAST FURNACE OZONATED WASTEWATER QUALITY

	Date			
	1/13	1/17	1/20	1/22
Run number	1	2	3	4
Contact time, min.	240	120	171	120
O ₃ applied, mg/l	1,054	1,205	1,074	739
O ₃ utilized, mg/l	363	364	480	612
Parameter				
pH	7.90	-	6.90	7.60
SS, mg/l	1	7	15	16
VSS, mg/l	<1	3	5	5
Phenol, mg/l	0.010	-	0.005	0.190
NH ₃ as N, mg/l	3.1	9.4	3.4	10
NO ₃ as N, mg/l	-	34.5	-	-
TKN as N, mg/l	1.1	4.8	0.67	12.7
S, mg/l	<0.2	<0.2	<0.2	0.1
CN, mg/l	-	<0.01	<0.01	0.03
CN _A , mg/l	<0.01	<0.01	<0.01	0.03
F, mg/l	10.8	11	12	16.2
TOC, mg/l	<2	6	5	1
BOD, mg/l	-	<5	-	-
Color, APHA units	5/10	0/5	5	10
O&G, mg/l	<1	-	-	-

Cost Data

Listed in Table 84 are the costs associated with the ozonation treatment train. The system has been designed to upgrade the blast furnace scrubber blowdown to meet BAT effluent limitations. Cost estimates are presented for design flows of 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd) and 5,678 m³/day (1.5 mgd).

TABLE 84. OZONATION TREATMENT TRAIN COST SUMMARY

Size, m ³ /day (mgd)	946 (0.25)	2,839 (0.75)	5,678 (1.5)
<u>Capital Costs (\$)</u>			
Filtration	\$ 195,000	\$ 355,000	\$ 533,000
Ozonation	4,727,000	12,793,000	24,415,000
Clarification	201,000	341,000	433,000
Total	\$5,123,000	\$13,489,000	\$25,381,000
<u>Operating Costs (\$/Yr)</u>			
Filtration	\$ 45,000	\$ 16,000	\$ 95,000
Ozonation	997,000	2,621,000	4,980,000
Clarification	35,000	62,000	84,000
Total	\$1,077,000	\$ 2,699,000	\$ 5,159,000
<u>Operating Costs, \$/3785 $\pm$ (\$/1000 gal.)</u>			
Filtration	0.49	0.24	0.17
Ozonation	10.92	9.58	9.09
Clarification	0.38	0.23	0.15
Total	11.79	10.05	9.41

The costs presented in Table 84 include pretreatment of the wastewater by dual media pressure filtration. The design filtration rate was assumed to be 203.7 l/min/m² (5 gpm/ft²). The ozone contactors were designed for a 60 minute contact time with an applied ozone dose of 733 mg/l. The cost data assume an air feed ozone system. The clarification system has been costed for an overflow rate of 22 m³/m²/day (530 gpm/ft²). Polyelectrolyte addition has been assumed.

Space Requirements

Space requirements for the ozonation treatment train are summarized below in Table 85. The area required includes the land necessary for filtration, ozonation and clarification. Sufficient space has been allowed for walkways, piping, etc. between the individual processes.

TABLE 85. OZONATION TREATMENT
TRAIN SPACE REQUIREMENTS

System capacity m ³ /day (mgd)	<u>946 (0.25)</u>	<u>2,839 (0.75)</u>	<u>5,678 (1.5)</u>
Space Requirements m ² (ft ²)			
Filtration	68 (700)	111 (1,200)	172 (1,800)
Ozonation	34 (360)	102 (1,100)	204 (2,200)
Clarification	<u>50 (540)</u>	<u>130 (1,400)</u>	<u>300 (3,100)</u>
Total	<u>152 (1,600)</u>	<u>343 (3,700)</u>	<u>676 (7,100)</u>

REVERSE OSMOSIS TREATMENT TRAIN

The reverse osmosis treatment train was the third successful group of process elements investigated for upgrading the BPT blast furnace scrubber blowdown to BAT discharge limitations. Technologies belonging to this treatment train included chemical treatment, filtration and reverse osmosis. Treatment of the brine was performed using either ozonation or alkaline chlorination.

Performance Data

Chemical treatment and clarification was required to remove suspended solids and turbidity from the scrubber blowdown prior to filtration and application to the reverse osmosis system. Chemical addition consisted of alum (50-150 mg/l) and polymer (0.5 mg/l). Turbidity removal averaged 68 percent while an 83 percent reduction in suspended solids was also achieved.

Dual media and magnetic filtration were the technologies investigated to provide final pretreatment to the wastewater before desalting by reverse osmosis. Superior suspended solids and turbidity removal was achieved by the dual media filter and this technology was used to provide feed to the reverse osmosis system.

The reverse osmosis testing program was performed in two phases. Phase 1 was operated at a 33 percent product water recovery while a 75 percent recovery was utilized in Phase 2. The two phases were used to obtain expected product water quality data and also cleaning information. TFC Model 4600 polyamide spiral wound membranes were used during the study.

The system was operated continuously over the test period and a total of 409.5 hours were put on the RO membranes.

Projected product water quality, assuming a 75 percent recovery is shown in

Table 86. The RO rejected significant percentages of all BAT parameters to the extent that the product water could be discharged directly or reused in the mill. Brine treatment by either alkaline chlorination or ozonation would be required prior to discharge. Information on treatment of this sidestream was also obtained during the study.

TABLE 86. BLAST FURNACE PROJECTED REVERSE OSMOSIS
PRODUCT QUALITY

Parameter	Product quality range
pH	5.35-5.75
SS, mg/l	3-4
VSS, mg/l	1-3
Phenol, mg/l	<0.054
NH ₃ as N, mg/l	9-10
NO ₃ as N, mg/l	<0.02
TKN as N, mg/l	10-12
S, mg/l	<0.2
CN, mg/l	<0.01
CN _A , mg/l	<0.01
F, mg/l	0.40-0.45
TOC, mg/l	<2
BOD ₅ , mg/l	<2
COD, mg/l	<5
TP as P, mg/l	<0.01
TDS, mg/l	60-70

Alkaline chlorination of the reverse osmosis brine consistently reduced the ammonia concentration to less than BAT levels, provided the chlorine to ammonia ratio was in excess of 7.6:1. Ammonia was found to be the limiting parameter. That is, if ammonia was removed to below BAT levels so were all other oxidizable parameters. Ammonia concentrations continued to drop as the Cl₂:NH₃ ratio was increased above 7.6:1 until breakpoint was reached at a ratio of about 10:1. Total cyanide and cyanide-A values were typically less than 0.01 mg/l for the effluent wastewater.

During ozonation of the reverse osmosis brine, the influent pH was elevated to 11.5-12.0 prior to applying ozone dosages ranging from 0-3,065 mg/l. Hydraulic detention times of 120-400 minutes were studied. To meet discharge requirements, a dosage of approximately 9 mg of ozone was required for each milligram of ammonia in the effluent.

Cost Data

The vendor prepared cost data for the reverse osmosis treatment train are shown in Tables 87 and 88. Cost estimates are listed for flows of 946 m³/day (0.25 mgd), 2,839 m³/day (0.75 mgd) and 5,678 m³/day (1.5 mgd). Table 87 assumes ozonation of the reverse osmosis brine, while the Table 88 cost estimates are for alkaline chlorination brine treatment. Again, unless dechlorination is required, activated carbon is not an essential part of the treatment system.

The pretreatment and post-treatment clarification systems assume an overflow rate of 22 m³/m²/day (530 gpd/ft²). The pretreatment pressure filtration unit is designed for a flow rate of 203.7 l/min/m² (5 gpm/ft²). The full-scale reverse osmosis unit assumes a product water recovery of 75 percent. The cost estimates for the reverse osmosis brine ozonation system are based on a design applied ozone dosage of 708 mg/l. For alkaline chlorination of the brine, a design chlorine to ammonia ratio of 10:1 using chlorine gas was assumed.

Space Requirements

The space requirements for selected reverse osmosis design flows are presented in Tables 89 and 90. Ozonation treatment of the brine is assumed in Table 89, while brine chlorination is assumed in Table 90. Comparison of the tables indicates that ozonation of the brine will result in a slight space savings.

DISCUSSION

The pilot program indicated that the alkaline chlorination, ozonation and reverse osmosis treatment trains were all capable of producing an effluent which would meet 1974 BAT guidelines. By reviewing the three evaluating criteria (performance, cost and space) for each treatment train, it is possible to choose an optimum system for final application. This has been addressed below:

1. With regard to performance, each of the three treatment trains investigated produced an effluent suitable for discharge.
2. Capital and operating costs are lowest for alkaline chlorination, independent of design flow. Ozonation is the least economical treatment train, requiring an expenditure of about twice that for alkaline chlorination.
3. The ozonation treatment train would require the least amount of installed equipment space. Alkaline chlorination and reverse osmosis space requirements were approximately equal, being about 30 percent greater than ozonation.

Using these criteria, conclusions can be drawn regarding the most attractive treatment process. Effluent water quality is essentially the same for the

TABLE 87. REVERSE OSMOSIS TREATMENT TRAIN COST SUMMARY
(assuming ozonation of the brine)

Size, m ³ /day (mgd) ^a	946 (0.25)	2,839 (0.75)	5,678 (1.5)
Capital Costs (\$)			
Clarification	\$ 201,000	\$ 341,000	\$ 433,000
Filtration	195,000	355,000	533,000
Reverse Osmosis	1,931,000	3,365,000	5,970,000
Ozonation-Brine	2,850,000	3,666,000	7,079,000
Clarification-Brine	143,000	193,000	257,000
Total	\$5,320,000	\$7,920,000	\$14,272,000
Operating Costs (\$/Yr)			
Clarification	\$ 38,000	\$ 70,000	\$ 101,000
Filtration	45,000	66,000	95,000
Reverse Osmosis	322,000	576,000	1,015,000
Ozonation-Brine	411,000	751,000	1,461,000
Clarification-Brine	28,000	35,000	44,000
Total	\$ 844,000	\$1,498,000	\$ 2,716,000
Operating Costs, \$/3,785 l (\$/1,000 gal.)^b			
Clarification	0.42	0.26	0.18
Filtration	0.49	0.24	0.17
Reverse Osmosis	3.53	2.10	1.85
Ozonation-Brine	4.50	2.74	2.67
Clarification-Brine	0.31	0.13	0.08
Total	9.25	5.47	4.95

a. System Feedwater

b. \$/3,785 l (\$/1,000 gal.) of system feedwater
[946, 2,839, 5,678 m³/day]

TABLE 88. REVERSE OSMOSIS TREATMENT TRAIN COST SUMMARY
(assuming chlorination of the brine)

Size, m ³ /day (mgd) ^a	946 (0.25)	2,839 (0.75)	5,678 (1.5)
<u>Capital Costs (\$)</u>			
Clarification	\$ 201,000	\$ 341,000	\$ 433,000
Filtration	195,000	355,000	533,000
Reverse Osmosis	1,931,000	3,365,000	5,970,000
Chlorination-Brine	384,000	457,000	507,000
Clarification-Brine	143,000	193,000	257,000
Total w/o Activated Carbon	\$2,854,000	\$4,711,000	\$7,700,000
Activated Carbon-Brine	807,000	945,000	1,320,000
Total w/Activated Carbon	\$3,661,000	\$5,656,000	\$9,020,000
<u>Operating Costs (\$/Yr)</u>			
Clarification	\$ 38,000	\$ 70,000	\$ 101,000
Filtration	45,000	66,000	95,000
Reverse Osmosis	322,000	576,000	1,015,000
Chlorination-Brine	204,000	432,000	591,000
Clarification-Brine	28,000	35,000	44,000
Total w/o Activated Carbon	\$ 637,000	\$1,179,000	\$1,846,000
Activated Carbon-Brine	175,000	292,000	491,000
Total w/Activated Carbon	\$ 812,000	\$1,471,000	\$2,337,000
<u>Operating Costs, \$/3,785 & (\$/1,000 gal.)^b</u>			
Clarification	0.42	0.26	0.18
Filtration	0.49	0.24	0.17
Reverse Osmosis	3.53	2.10	1.85
Chlorination-Brine	2.24	1.58	1.08
Clarification-Brine	0.31	0.13	0.08
Total w/o Activated Carbon	6.99	4.31	3.36
Activated Carbon-Brine	1.92	1.07	0.90
Total w/Activated Carbon	8.91	5.38	4.26

a. System Feedwater

b. \$/3,785 & (\$/1,000 gal.) of system feedwater
[946, 2,839, 5,678 m³/day]

TABLE 89. REVERSE OSMOSIS TREATMENT TRAIN SPACE REQUIREMENTS
(assuming ozonation of the brine)

Size, m ³ /day (mgd)	<u>946 (0.25)</u>	<u>2,839 (0.75)</u>	<u>5,678 (1.5)</u>
Space Requirements m ² (ft ²)			
Clarification	50 (450)	130 (1,400)	300 (3,100)
Filtration	68 (700)	111 (1,200)	172 (1,800)
Reverse Osmosis	71 (750)	167 (1,800)	333 (3,600)
Brine Ozonation	41 (450)	62 (670)	94 (1,000)
Brine Clarification	12 (120)	30 (350)	70 (780)
Total	<u>242 (2,560)</u>	<u>500 (5,420)</u>	<u>969 (10,280)</u>

TABLE 90. REVERSE OSMOSIS TREATMENT TRAIN SPACE REQUIREMENTS
(assuming chlorination of the brine)

Size, m ³ /day (mgd)	<u>946 (0.25)</u>	<u>2,839 (0.75)</u>	<u>5,678 (1.5)</u>
Space Requirements m ² (ft ²)			
Clarification	50 (540)	130 (1,400)	300 (3,100)
Filtration	68 (700)	111 (1,200)	172 (1,800)
Reverse Osmosis	71 (750)	107 (1,800)	333 (3,600)
Brine Chlorination	139 (1,500)	153 (1,650)	179 (1,920)
Brine Clarification	12 (120)	30 (350)	70 (780)
Brine-Act. Carbon	223 (2,400)	241 (2,600)	265 (2,850)
Total	<u>563 (6,010)</u>	<u>832 (9,000)</u>	<u>1,319 (14,050)</u>

three treatment trains. Therefore, the selection of the most feasible process for this application should be based on the two remaining criteria, namely cost and space. Using these guidelines, it is obvious that alkaline chlorination is the most cost-effective treatment train and would be recommended as the optimum process unless severe space requirements exist at the site. If economy of space is the limiting criterion, the study results have indicated that ozonation would be the preferred process.

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APPENDIX A-1
BENCH TEST FINDINGS

INTRODUCTION

Bench scale treatment tests were performed by the Environmental Research Center of Rexnord on a sample of blast furnace clarifier effluent from a Midwestern steel plant. The laboratory bench scale tests were performed in the Rexnord laboratory from March 15-30, 1978. The purpose of these tests was to investigate treatment feasibility; obtain basic design data for the pilot plant; determine waste pretreatment requirements; and generate the information required to plan the pilot plant tests. The bench scale processes investigated were clarification, chlorination, filtration, carbon adsorption, ozonation, and reverse osmosis.

The bench tests were performed to investigate treatment methods for meeting BAT effluent requirements for the blast furnace-iron subcategory and ferro-manganese subcategories. The target effluent concentrations are defined as follows:

<u>Parameter</u>	<u>Target effluent concentration to meet BAT requirement</u>
pH	6-9
Ammonia	<10 mg/l
Cyanide	<0.25 mg/l
Fluoride	<20 mg/l
Phenol	<0.5 mg/l
Sulfide	<0.3 mg/l
Suspended Solids	<10 mg/l
Manganese	<5 mg/l

SAMPLE ACQUISITION

Samples were procured from the blast furnace clarifier effluent launder of a Midwestern steel plant at about 11:00 a.m. on March 15, 1978. This was the most accessible location to obtain a representative sample. Approximately 833 liters (220 gallons) of waste was filled into four 208 liters (55 gal.) drums. A schematic of the sampling location is shown in Figure A-1.1. At the time of sampling, blast furnace No. 1 was not operating due to repair (mini-relining). The blast furnace No. 1 clarifier waste was being pumped to blast furnace No. 2 clarifier at a rate of about 757 lpm (200 gpm).

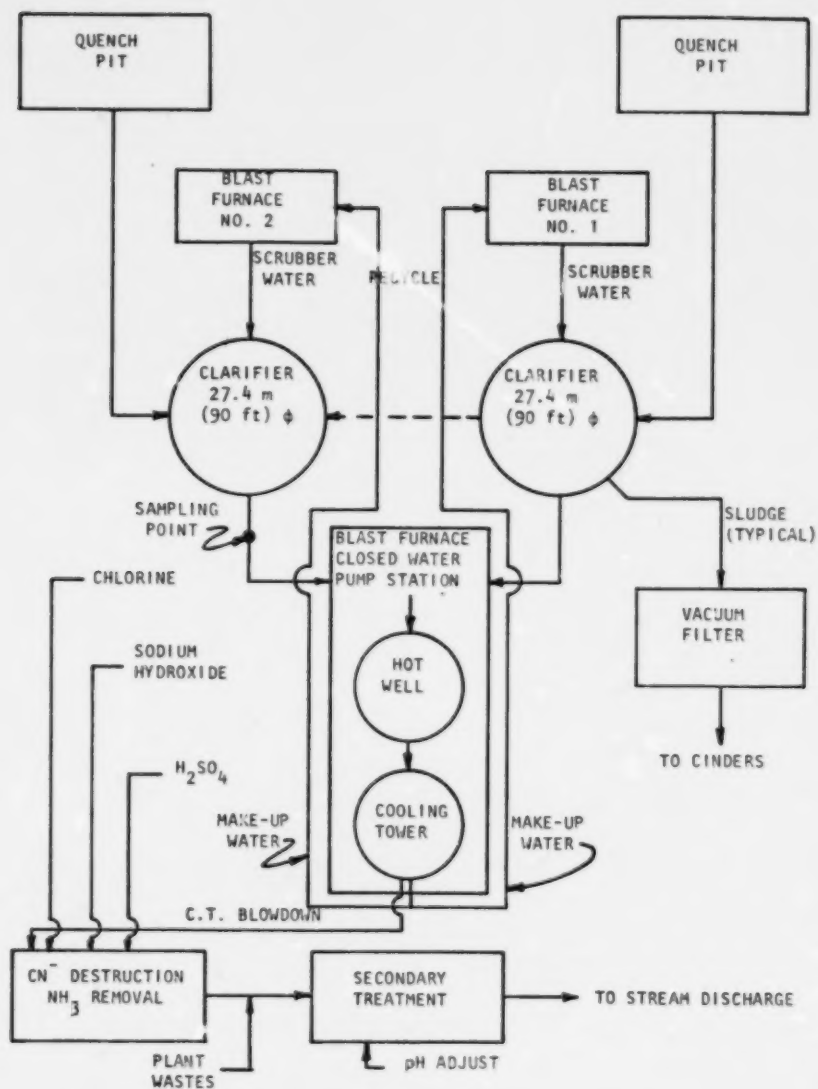


Figure A-1.1. Blast furnace effluent sampling location.

Typically, interpumping between clarifiers is not conducted.

After acquisition, the wastewater was transported to the Rexnord laboratory by truck. The drums were refrigerated from the time of arrival at the laboratory (6:00 p.m., 3/15/78) until the bench testing was completed.

EXPERIMENTAL EQUIPMENT AND PROCEDURES

Dual Media Filter

The laboratory dual media filter (DMF) was constructed from a 8.26 cm ID x 106.7 cm long (3.25 in. x 42 in.) plexiglas tube. A fine mesh screen was used at the bottom to support the filter media. Rubber stoppers were used to seal each end of the tube. Fittings were installed in the stoppers to allow downflow for filtration or upflow for backwash. The DMF effluent fitting at the bottom of the tube was connected by rubber hose to a head control box attached to the side of the column above the level of the filter media. This box ensured that the filter media would remain submerged when flow to the DMF was reduced or stopped. A garden hose connection at the bottom of the column allowed the use of tap water for backwashing the filter.

Red Flint filter sand from Carpenter Brothers Inc. (Milwaukee, WI) was used as the support sand. The anthracite coal was obtained from Reading Anthracite Coal Company, Pottsville, Pennsylvania. To prepare the column, sand was added to a depth of 22.9 cm (9 in.) followed by coal to a depth of 45.7 cm (18 in.) for a total filter depth of 68.6 cm (27 in.). The filter was backwashed with tap water until the backwash water was clear. Waste was pumped to the filter with a variable speed Randolph pump (The Randolph Company, Houston, Texas). Flows were measured by the pail and stopwatch method.

Reverse Osmosis Equipment

The Rexnord laboratory reverse osmosis unit was used for these tests. This unit utilizes a 10.2 cm (4 in.) DuPont reverse osmosis module containing a Permasep B-9 hollow fiber membrane. A Cat pump was used to supply the feed to the reverse osmosis module at pressures up to 28 kg/cm² gage (400 psig). A ten micron cartridge filter was installed upstream of the Cat pump to remove large particles that may be harmful to the pump or to the hollow fiber module. Flow to the entire system was supplied by a centrifugal pump acting as a booster pump.

Ozone Test Apparatus and Procedure

Ozone for the study was supplied from an ozonator manufactured by the Ozone Research and Equipment Corporation (OREC). The model number was 03B1-0. Oxygen was used as the feed gas.

During the bench study, two treatment techniques were used. For the low

dosage tests, ozone was contacted with the wastewater using an 1100 cc ozone reaction cell prepared for Rexnord by Pope Scientific Co. (Menomonee Falls, WI). A sketch of the tube is shown in Figure A-1.2. The reaction cell was filled by immersing inlet C in the wastewater and applying a vacuum at tube A. The stopcocks were closed when the waste entered tube A. The tubing containing gaseous ozone was connected to tube B and the stopcocks opened so that the ozone could enter the cell as a portion of the waste was drained through tube D to a graduated cylinder. The volume of waste drained from the cell was used to measure the volume of gaseous ozone added to the cell. The volume of ozone gas added to the cell ranged from 46 cc to 515 cc in these tests. After the addition of the ozone, the stopcocks were closed and the cell was shaken for the desired period of time. The treated waste was then drained from the ozone cell for analysis.

The concentration of ozone in the gas flow from the ozonator was measured before and after each test. For these standardizations, a 275 cc ozone reaction cell was used. The cell was filled with 2 percent KI solution and ozone gas was added in the same manner as ozone was added to the wastewater flask. After reaction of the ozone and the KI by shaking for two minutes, the iodine formed was measured by iodometric titration. There was little change in ozone concentration between tests. In one series of tests using this apparatus, the ozonator gas averaged 44.6 mg O_3/l gas and in a second series of tests, the gas averaged 55.5 mg O_3/l gas.

The oxidation of ammonia by ozone requires a high dosage of ozone. Due to the size limitation of the reaction cell, a different treatment technique was used for this phase of the testing.

For the high dosage tests, ozone produced from the OREC generator was fed to a series of four reaction flasks. Flasks No. 1 and 2 each contained 500 ml of wastewater. Flasks No. 3 and 4 each contained 200 ml of a 2 percent potassium iodide solution which served as a trap to catch the excess ozone passing through the waste. During the ozonation period, flasks No. 1 and 2 were switched midway through the test so that each waste sample was first in the train for half of the time. The contents of flasks No. 1 and 2 were combined for analyses. All tests were run at room temperature.

The ozone dosage applied to the wastewater in these tests was controlled by varying the amount of time that the waste was ozonated. Gas from the ozonator was fed to the absorption flasks at a steady rate of about 700 cc/min.. Before and after each ozonation test, 1 percent KI solution was placed in flask No. 1 and ozonated for a measured period to time. The amount of ozone absorbed was determined by iodometric titration and the rate of ozone addition to the system determined. The wastewater and KI solutions in flasks 2, 3 and 4 were replaced with identical volumes of water so that the flow through the system was identical to that during the ozonation tests. During ozonation of the wastewater, the amount of ozone trapped in flasks 3 and 4 was subtracted from the amount added to the system to obtain the amount absorbed by the wastewater. In preliminary standardization tests using KI solution in the first two flasks, it was

4.7 cm



20 cm

3.45 cm

Note: $\text{cm} \times 0.39 = \text{inches}$

Figure A-1.2. Ozone reaction bulb with 3-way teflon stopcocks.

determined that 99.9 percent of the ozone was absorbed in the first flask. During ozonation of wastewater, flask No. 3 trapped 99.9 percent of the ozone trapped by flasks 3 and 4.

All parts in contact with the ozone were made of stainless steel, Teflon, glass or new Tygon. All parts were contacted with ozone prior to testing.

Chlorination Test Procedure

Chlorine oxidation tests were performed on the raw wastewater to determine its effectiveness in removing cyanide and ammonia. The wastewater was adjusted to a pH of 10.7 ± 0.2 with sodium hydroxide and the chlorine in the form of sodium hypochlorite was added. The wastewater was held for 7 minutes, then the pH was lowered to 7.4 ± 0.2 with hydrochloric acid. The detention time was based on literature values (21,22). The batch was flocculated with 50 mg/l FeCl_3 and 0.5 mg/l Betz 1100 polyelectrolyte and allowed to settle for 30 minutes. Residual chlorine was measured immediately on samples withdrawn from the supernatant. Excess chlorine was removed from the ammonia and cyanide samples with ascorbic acid and the samples were preserved for later analysis.

Carbon Adsorption Test Equipment and Procedure

Pulverized activated carbon was obtained directly from the manufacturers for use in these tests. Weights of carbon ranging from about 0.01 g (.0004 oz.) to about 4.5 g (.16 oz) were placed in seven flasks for each carbon isotherm test. Tests were performed using a 300 ml (.08 gal.) sample per flask. In addition, an eighth flask containing sample but no carbon was used as a blank. After shaking for 30 minutes using a Burrell Wrist-Action Shaker, the contents of the eight flasks were filtered through washed 0.45 μm membrane filters to remove suspended material. The filtrate was saved for analysis.

Analytical Procedures

The following analytical procedures were used. In this listing, Standard Methods refers to Standard Methods for the Examination of Water and Wastewater, 14th Ed., APHA, AWWA, and WPCF, 1975 (2).

1. pH - Beckman Electromate pH meter, Standard Methods, p. 460.
2. Turbidity - Hach model 2100A turbidimeter, Standard Methods, p. 132.
3. Color - Hellige Aquatester with color comparison discs calibrated at 0, 5, 10, 20, 30, 40, 50, 60 and 70 APHA color units. Waste samples were filtered and diluted with distilled water (if necessary) to reduce color to less than 70 APHA units. When the color of the sample was clearly between that of two color discs, the values for both color discs were recorded. For example, the term 0/5 means that the color of the sample was more than 0 APHA units but less than 5 APHA units. For plotting and calculation purposes, the value mid-way between the two color discs was used.
4. BOD - Standard Methods, p. 543.

5. Alkalinity - Standard Methods, p. 278.
6. Total Organic Carbon (TOC) - Beckman Model 915A TOC Analyzer Standard Methods, p. 532.
7. Suspended Solids - Standard Methods, p. 94.
8. Oil and Grease - Standard Methods, p. 515.
9. Total Phosphate - Standard Methods, p. 476 and p. 624.
10. Total Dissolved Solids - Standard Methods, p. 92.
11. Ca and Total Hardness - Standard Methods, p. 202.
12. Sulfate - Standard Methods, p. 496.
13. Chloride - Standard Methods, p. 304.
14. Conductivity - Standard Methods, p. 71.
15. Nitrate Nitrogen - Standard Methods, p. 437.
16. Metals - Na, K, Mn, Fe, Al, Ca, Mg - Standard Methods, p. 144.
17. Ammonia, Orion Ion Specific Electrode - EPA Methods for Chemical Analysis of Water and Wastes, 1974, p. 165.
18. Cyanide - Distillation followed by Orion Ion Specific Electrode, Standard Methods, p. 367.
19. Residual Chlorine - Iodometric Titration, Standard Methods, p. 318.
20. Phenol - Distillation followed by 4-AAP colorimetric method, Standard Methods, p. 576.

RAW WASTE CHARACTERISTICS

Comparison of BPT and BAT Effluent Limitations and Raw Waste

The comparison between the waste sample analytical values and the Iron and Steel blast furnace subcategory BPT and BAT limitations are shown in Table 1. For those parameters for which BPT limitations have been set (namely ammonia, cyanide, phenol, sulfide, suspended solids, and pH), none were exceeded by the waste sample. For most of the raw waste sample parameters, analytical values are well below those listed for blast furnace BPT effluent. A listing of additional analyses performed on the waste are also presented in Table A-1.1.

Table A-1.1 also shows that the waste sample has a higher contaminant concentration for many of the guideline parameters listed than that allowed for BAT requirements. This indicates that further treatment will be necessary to meet future federal BAT effluent goals. The objective of this bench testing program was to investigate various advanced treatment technologies for meeting blast furnace BAT limitations. The results and recommendations of the bench study are presented in the following sections.

CLARIFIER EFFLUENT PRETREATMENT

Preliminary Treatment Tests

A representative sample of wastewater was obtained from the four barrels after arrival in the Rexnord laboratory to perform preliminary treatment

TABLE A-1.1. RAW WASTEWATER ANALYTICAL
VALUES INCLUDING BPT AND BAT EFFLUENT LIMITATIONS
FOR BLAST FURNACE OPERATIONS (23)

Parameter	Units	BPTCA Limitations(1)		Raw waste sample analytical values	BATEA Limitations(1)	
		Blast furnace- ferromanganese subcategory	Blast furnace- iron subcategory		Blast furnace- ferromanganese subcategory	Blast furnace- iron subcategory
Ammonia	mg/l	<200	<125	20.1	<10	<10
Cyanide	mg/l	<30	<15	0.64	<0.25	<0.25
Manganese	mg/l	<16	-	2.8	<5	-
Phenol	mg/l	<4	<4	0.045	<0.5	<0.5
Sulfide	mg/l	<30	<6	<0.1	<0.3	<0.3
Suspended solids	mg/l	<100	<50	34	<25	<10
pH	-	6-9	6-9	7.8	6-9	6-9
Fluoride	mg/l	-	<40	8.9	-	<20
Total dissolved solids	mg/l	-	-	712	-	-
Turbidity	FTU	-	-	35	-	-
Alkalinity	mg/l as CaCO ₃	-	-	210	-	-
BOD	mg/l	-	-	6	-	-
TOC	mg/l	-	-	12	-	-
Oil and grease	mg/l	-	-	<1	-	-
Chloride	mg/l	-	-	220	-	-
Sulfate	mg/l	-	-	70	-	-
Sodium	mg/l	-	-	23	-	-
Potassium	mg/l	-	-	14	-	-
Calcium	mg/l	-	-	120	-	-
Magnesium	mg/l	-	-	34	-	-
Iron	mg/l	-	-	0.51	-	-

tests. One of the advanced technologies to be investigated in the pilot test program (reverse osmosis) requires a high clarity wastewater to prevent fouling of the equipment. A feed turbidity of less than 1 FTU is recommended for use of a spiral wound reverse osmosis module. Therefore, chemical pretreatment jar tests were performed using alum and ferric salts.

The effect of various alum treatment conditions on turbidity is shown in Table A-1.2. The treatment train consisted of alum addition, a 10 second rapid mix period, polymer addition, and a 2.5 minute flocculation period. This procedure was used for all but one test. These conditions produced maximum clarity. These results show that:

1. Alum dosages from 50 to 150 mg/l were capable of producing a wastewater turbidity of less than 1 FTU.
2. The effluents were filtered through a Whatman No. 541 paper prior to taking the turbidity readings. Blinding of this paper occurred quite rapidly indicating that a dual media filter could experience rapid headloss.
3. Settling rates of the flocculated solids were less than 30.5 cm/min (1 ft/min) with a large amount of lagging fines after 30 minutes of settling.

Ferric chloride jar tests were also performed on the wastewater. Optimum clarity conditions were produced using a 10 second rapid mix followed by a 2.5 min. flocculation time. Organic polyelectrolyte was added to enhance floc formation. The test results are presented in Table A-1.3. The ferric chloride jar testing data indicate that:

1. Ferric chloride was capable of reducing effluent turbidity to less than 1 FTU using dosages of 50-75 mg/l.
2. Mix times longer than 10 seconds did not improve effluent clarity.
3. A maximum flocculation time of 2.5 minutes was necessary for minimum effluent turbidity.
4. Ferric chloride produced a clearer supernatant and a more rapid settling floc than alum.
5. Acidifying the effluent from Test No. 7 to pH 6.0 increased the turbidity to 0.85 FTU.

Gravity Separation Tests

Gravity separation tests were performed using ferric chloride and polymer addition. The chemical addition procedure, settling data, and effluent quality results are presented in Table A-1.4. The detention time for Test No. 1 was 30 minutes; for Test No. 2, a detention time of 60 minutes was used.

A settling rate of 7.3 cm/min (0.24 ft/min) was observed for both tests. Sludge volumes were also equivalent for the 30 minute and 60 minute tests (8 ml/liter [8 gal./1000 gal.]). Effluent quality data indicated that there was no advantage to increasing the settling time from 30 minutes to 60 minutes. Turbidity in both tests was 2.2 FTU's which is above the 1 FTU

TABLE A-1.2. RESULTS OF PRELIMINARY ALUM
TREATMENT TESTS

Test No.	1	2	3	4	5	6	7	8
Chemical Treatment: $\text{Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}$, mg/ℓ	150	150	150	150	150	150	100	50
Rapid Mix, sec.	10	10	10	10	10	10	10	10
Percol 727, mg/ℓ	-	-	-	0.5	0.75	-	-	-
Percol 730, mg/ℓ	0.5	0.75	1.0	-	-	0.75	.75	0.75
Floc Time, min.	2.5	2.5	2.5	2.5	2.5	1.0	2.5	2.5
Feed Quality: pH: 7.8								
Turbidity, FTU: 35								
Effluent Quality: pH	6.5	6.5	6.5	6.5	6.5	6.5	6.7	6.9
Turbidity, FTU	0.86	0.80	1.10	1.20	1.25	1.4	0.75	0.72

TABLE A-1.3. RESULTS OF PRELIMINARY FERRIC
CHLORIDE TREATMENT TESTS

Test No.	1	2	3	4	5	6	7	8	9
Chemical Treatment:									
FeCl ₃ , mg/ℓ	100	75	50	50	25	50	50	50	50
Rapid mix, sec.	10	10	10	10	10	10	10	10	10
Percol 730, mg/ℓ	1	1	1	2	0.75	-	-	-	-
Percol 727, mg/ℓ	-	-	-	-	-	1	-	-	-
Betz 1100, mg/ℓ	-	-	-	-	-	-	0.5	0.75	1
Floc time, min.	2.5	2.5	2.5	2.5	2.5	2.5	2.5	2.5	2.5
Feed Quality:									
pH: 7.8									
Turbidity, FTU: 35									
Iron, mg/ℓ: 0.51									
Effluent Quality:									
pH	6.3	6.5	6.7	6.7	7.1	6.7	6.7	6.7	6.7
Turbidity, FTU's	1.4	0.9	0.7	1.3	2.1	1.0	0.6	1.2	1.2
Iron, mg/ℓ	0.55	0.55	0.50	-	0.85	-	-	-	-

TABLE A-1.4. RESULTS OF RAW WASTE SETTLING TESTS

Test No.	1	2
<u>Chemical Treatment:</u>		
FeCl ₃ , mg/l	50	50
Rapid mix, sec.	10	10
Betz 1100, mg/l	0.5	0.5
Floc time, min.	2.5	2.5
<u>Settling Data:</u>		
Settling rate, cm/min (fpm)	7.3 (0.24)	7.3 (0.24)
Detention time, min.	30	60
Sludge vol., ml/l (gal./10 ³ gal.)	8	8
Scum vol., ml/l (gal./10 ³ gal.)	Trace	Trace
<u>Effluent Quality:</u>		
pH	7.0	7.0
Turbidity, FTU's	2.2	2.2
SS, mg/l	7	8
Alk., mg/l	150	155
TOC, mg/l	11	11
BOD, mg/l	11	10
Chloride, mg/l	255	250
Fe, mg/l	1.05	1.20
TDS, mg/l	734	722
Fluoride, mg/l	7.9	7.8
Silica, mg/l as Si	6	6

value recommended for operation of a reverse osmosis spiral wound module.

Recommendations for Pilot Test Program

It is recommended that the wastewater be chemically treated with ferric chloride and polyelectrolyte and clarified by settling. Ferric chloride dosages of 50 mg/l and a Betz 1100 dosage of 0.5 mg/l can be used to plan preliminary chemical feed concentrations and feed rates. It is expected that the required dosage may vary as the wastewater characteristics vary and it is recommended that the optimum chemical dosage be verified during the pilot tests.

Pilot tests should also investigate the use of alum if turbidity requirements cannot be met using ferric salts. Dosages of 50-150 mg/l [as Al₂(SO₄)₃ · 18H₂O] and a Percol 730 polymer addition of 0.75 mg/l could serve as a suitable alternate to the ferric chloride/Betz polymer treatment.

DUAL MEDIA FILTRATION

Dual media filtration will be used in the pilot testing as a polishing step and to reduce the suspended solids concentration of the wastewater prior to treatment by ozone, activated carbon and reverse osmosis. Similarly, a laboratory dual media filter was used in the bench tests to provide a low-solids waste for the ozone, carbon isotherm and reverse osmosis tests.

Red Flint filter sand having a nominal grain size of 0.5 mm was used for all the bench tests. The coal size tested was 1.0-1.1 mm. In all tests, sand depth was 22.2 cm (8.75 in.) and coal depth was 45 cm (17.75 in.).

Barrel Treatment - Raw Waste

A 151.4 liters (40 gal.) batch of raw wastewater was passed through the dual media filter at an average rate of 148.7 lpm/m^2 (3.65 gpm/ft^2). Samples were taken after 30 minutes and 60 minutes. The data presented in Table A-1.5 indicate that turbidity and suspended solids concentrations were reduced with length of filter run. Oil and grease, TOC, and pH were unchanged. After a 60 minute run time, turbidity was 4.0 FTU which would be unacceptable for use with a reverse osmosis unit due to rapid fouling of the membranes. There was no substantial headloss variations between the 30 minute and 60 minute readings. Filter headloss at both of these times was 31.1 cm (12.25 in.) of water. The backwash was observed to contain black suspended solids.

Barrel Treatment - Chemically Pretreated Waste

Using the information obtained from the gravity separation tests, a 151.4 liters (40 gal.) sample of raw waste was treated with 50 mg/l FeCl_3 and mixed for 30 seconds. Polyelectrolyte (0.5 mg/l Betz 1100) was then added and the waste was flocculated for 2.5 minutes and then allowed to settle. After a one hour detention time, the supernatant was decanted below a layer of floating scum.

The clarified effluent was added to the dual media filter (sand/anthracite) at a rate of 150.7 lpm/m^2 (3.7 gpm/ft^2) for 85 minutes. Samples of effluent were taken after 40 minutes and 85 minutes for analyses. The results of these analyses are listed in Table A-1.6. The data indicate that turbidity improved with length of filter run. There was no significant increase in headloss between the 40 minute and 85 minute readings (30.5 cm [12 in.] of water). The filter backwash water contained a brown colored precipitate.

Extensive analytical data on the raw waste sample, the chemically treated effluent, and the dual media filter effluent are shown in Table A-1.7. As indicated by the data, turbidity following filtration was reduced to 0.4 FTU. Acidification to pH $\approx$ 6.5 prior to reverse osmosis treatment may result in a slight increase in turbidity, as observed previously.

Recommendations for Pilot Tests

It is recommended that the 1.0-1.1 mm (.039-.043 in.) anthracite coal be

TABLE A-1.5. DUAL MEDIA FILTRATION OF RAW WASTEWATER

Sample	Feed	Effluent	Effluent
Time after start, hr	-	0.5	1
Flow, lpm/m ² (gpm/ft ²)	-	147 (3.6)	151 (3.7)
Filter headloss, cm H ₂ O (in.)	-	31.1 (12.25)	31.1 (12.25)
Sample:			
pH	7.15	7.20	7.20
Turbidity, FTU	38	4.9	4.0
Suspended solids, mg/l	48	19	13
Oil and Grease, mg/l	<1	<1	<1
TOC, mg/l	14	8	8

TABLE A-1.6. DUAL MEDIA FILTRATION OF CHEMICALLY TREATED WASTE

Sample	Settling Test Effluent	DMF Effluent	DMF Effluent
Time after start, min.	-	40	85
Flow, lpm/m ² (gpm/ft ²)	-	151 (3.7)	151 (3.7)
Headloss, cm H ₂ O (in.)	-	30.5 (12)	30.5 (12)
Sample:			
pH	7.0	7.0	7.0
Turbidity, FTU	4.1	0.42	0.30 ^a
Suspended Solids, mg/l	10	3	-

a. After acidification to pH 6.4 with 44 mg/l HCl, turbidity increased to 0.38 FTU.

TABLE A-1.7. RESULTS OF ANALYSIS OF CLARIFIED WASTEWATER

	Composite raw sample	After Chem ^a treatment	After DMF (RO feed)
pH	7.8	7.0	6.55
Suspended solids, mg/l	34	7	-
Total dissolved solids, mg/l	712	734	740
DS meter reading, mg/l	-	-	760
Turbidity, FTU	35	2.2	0.4
Alkalinity, mg/l as CaCO ₃	210	150	95
BOD, mg/l	6	11	9
COD, mg/l	-	-	35
TOC, mg/l	12	11	7
Oil and Grease, mg/l	<1	-	-
Cyanide, mg/l	0.64	-	0.15
Phenol, mg/l	0.045	-	0.036
Fluoride, mg/l	8.9	7.9	6.7
NH ₃ , mg/l	20.1	-	13.3
TKN, mg/l	-	-	17
NO ₃ , mg/l	-	-	0.17
Chloride, mg/l	220	255	274
Sulfate, mg/l	70	-	70
Sulfide, mg/l	<0.1	-	-
Sodium, mg/l	23	-	20
Potassium, mg/l	14	-	9.7
Calcium, mg/l	120	-	115
Magnesium, mg/l	34	-	31.5
Manganese, mg/l	2.8	-	2.6
Iron, mg/l	0.51	1.05	0.11
Silica, mg/l	-	-	-

a. Chemical Treatment: 50 mg/l FeCl₃, 0.5 mg/l Betz 1100,
mix 30 sec., floc 2.5 min.

used in the pilot tests. This recommendation is based on previous test work (24) that indicated this size coal has superior backwashing characteristics to smaller or larger coal sizes. The bench data also indicate that chemical treatment followed by dual media filtration will reduce the turbidity of the waste to less than 1 FTU. A high clarity feed of less than 1 FTU turbidity level is recommended as the maximum for use of a spiral wound reverse osmosis module.

REVERSE OSMOSIS

The purpose of the reverse osmosis (RO) bench tests was 1) to determine the degree of pretreatment required to prevent fouling of the RO membrane and thereby enhance the feasibility of utilizing RO for treating the wastewater, and 2) to approximate RO operating conditions and the quality of the product and brine produced. The data obtained during the bench tests will be used to select the type of RO membranes, the pretreatment methods, and the operating conditions for the pilot study.

Pretreatment Evaluation

A relatively clear feed, to prevent fouling of the reverse osmosis membrane, is needed to utilize the RO treatment process. Generally, pretreatment is indicated when a wastewater contains a significant amount of suspended solids and/or the turbidity is greater than one FTU.

It is evident from the raw waste characteristics (suspended solids and turbidity from Table A-1.7) that pretreatment prior to RO is necessary. Moreover, it is further apparent from previous bench work performed on this wastewater and reported in another section that chemical treatment, sedimentation and filtration pretreatment is indicated prior to RO. The results of pretreatment of the raw wastewater by chemical treatment followed by filtration are presented in Table A-1.8 under the column labeled "Feed". Examination of the filtered wastewater characteristics indicate that the pretreatment used will eliminate gross fouling of the membranes.

The pretreated wastewater was further investigated to establish any additional membrane fouling potential. The parameters used to measure this potential included calcium carbonate and calcium sulfate precipitation and Millipore Pluggage Index (MPI). The investigation produced the following results to obtain a product water recovery of 75 percent (where product water recovery is the ratio of the product water volume to the feed volume):

1. To prevent calcium carbonate precipitation, the RO feed pH should be adjusted to about 6.4.
2. Calcium sulfate precipitation does not appear to be a problem up to product water recoveries of about 75 percent.
3. A MPI of 15 was observed.

Three membrane configurations are on the market, namely, tubular, spiral

TABLE A-1.8. REVERSE OSMOSIS TEST RESULTS

Parameter	Feed	Product	Brine
pH	6.55	6.4	7.7
Alkalinity, mg/l as CaCO_3	95	26	320
Dissolved solids, mg/l	740	90	2,980
Turbidity, FTU	0.42	0.18	0.48
BOD, mg/l	9	10	19
TOC, mg/l	7	5	30
COD, mg/l	35	16	95
Sulfate, mg/l	70	<1	285
Chloride, mg/l	274	36	1,090
Cyanide, mg/l	0.15	<0.1	0.80
Phenol, mg/l	0.036	0.012	0.18
Nitrate, mg/l as N	0.17	0.04	0.66
Ammonia, mg/l as N	13.3	3.9	40.1
Kjeldahl nitrogen, mg/l as N	17	6	61
Sodium, mg/l	20	5.5	51
Potassium, mg/l	9.7	3.0	39
Calcium, mg/l	115	8.1	475
Magnesium, mg/l	31.5	2.8	145
Iron, mg/l	0.11	0.04	0.39
Manganese, mg/l	2.6	0.11	9.4
Fluoride, mg/l	6.7	0.65	20.4
Silica, mg/l as SiO_2	13	-	-

Operating Conditions:

Feed: Dual media filter effluent after chemical treatment

Operating pressure: 274,209 kg/m^2 (390 psi)

Operating temperature: 20.5°C

Feed Flow: 7.7 lpm (2.03 gpm)

Product Flow: 6 lpm (1.58 gpm)

Brine Flow: 1.7 lpm (0.45 gpm)

Recovery: 78%

wound and hollow fiber. The most widely used membrane configurations at present utilize the spiral wound and hollow fiber membranes. The comparative disadvantages of the tubular configuration are appreciable and include higher expense (operating and capital) and the relatively small membrane surface area to unit volume ratio (resulting in more required space and low productivity). Further consideration to the tubular design will not be given for this application.

The results indicate that the quality of the feed after pretreatment is such that only the spiral wound RO design could be expected to operate satisfactorily. DuPont, a manufacturer of the hollow fiber type module, does not recommend operation unless the MPI is less than 3. To achieve this type of quality, pretreatment would have to be provided in addition to chemical treatment and filtration. The cost of this additional pretreatment would make application of the hollow fiber design impractical.

Reverse Osmosis Tests

The objective of the second portion of the bench testing work was to establish the approximate RO operating conditions and to determine the quality of the brine and product water produced. For this phase, a sample of wastewater was chemically treated and filtered through a dual media filter. The effluent was collected and then fed to the Rexnord portable laboratory RO unit.

The operating conditions at which the tests were run and the quality of the various streams are summarized in Table A-1.8. The tests indicated that 75 percent recovery is feasible. The brine showed no evidence of post-precipitation. Ammonia rejection was 71 percent at a feed concentration of 13.3 mg/l. Assuming the same rejection at higher initial ammonia concentrations, the initial feed ammonia concentration may be as much as 34 mg/l without the product concentration exceeding the BAT allowable 10 mg/l $\text{NH}_3\text{-N}$. The test results also show that the product water quality met the BAT requirements for cyanide, phenol, pH, fluoride and manganese.

Recommendations for Pilot Study

Reverse osmosis resulted in substantial reductions of all parameters tested and should be further evaluated during the pilot portion of this project. Chemical treatment followed by dual media filtration should be used as pretreatment before reverse osmosis. Because of the fouling potential of the wastewater, it is not practical to use the hollow fiber design during the pilot study. The spiral wound membrane configuration will be tested during the pilot study. The reverse osmosis system can be operated at 75 percent recovery.

OXONATION

Wastewater Treatment

Ozonation tests were performed to study the effect of ozone on cyanide,

phenol and ammonia concentrations of the wastewater and the reverse osmosis brine. The objective was to determine the ozone dosage required and the optimum conditions for use of ozone.

Destruction of phenol by ozonation was reported by Hardisty and Rosen (25) to be most efficient at pH greater than 9 with the maximum reaction rate occurring at pH 11.5. The amount of ozone needed was about 4.5 mg O_3 /mg phenol. Cleary and Kinney (26) also found that the optimum pH for ozonation of phenol was 11.8, but they could not correlate ozone dosage to phenol concentration when the wastewater (effluent from a coke plant dephenolized ammonia still) contained other oxidizable material. They found that temperature had no effect on ozone required and that the ammonia concentration, which varied widely, also did not affect ozone dosage. Hardisty and Rosen (25) also reported on the ozonation of simple and complex cyanides. They found that ozone dosages of 1.85 and 3.8 mg O_3 /mg CN were needed to oxidize simple and readily dissociated complex cyanides. Iron cyanide complexes, however, were much more difficult to destroy. Probes, Melnyk and Mansfield (27) found that a combination of ozone and ultra violet was much more effective than ozone alone in destroying the ferricyanide complex. Using the ozone-UV treatment, the rate of oxidation was better at pH 7 than at pH 4.5 or 9.5. They also found that the amount of ozone needed ranged from 23 mg O_3 /mg CN for a solution containing only ferricyanide to 308 mg O_3 /mg CN for a simulated blast furnace effluent containing 70 mg/l ammonia, 140 mg/l chloride, 0.2 mg/l phenol and 1 mg/l cyanide in the form of the ferricyanide. Hoigné and Bader (28) showed that the optimum pH for ozonation of ammonia was between 9 and 10. The presence of carbonate ions reduced the efficiency of oxidation at pH values greater than 9. At pH levels lower than 9, however, the rate of reaction was very slow due to the low concentration of free NH_3 . Singer and Zilli (29) found only a slight decrease in ammonia when ozonating at pH 7.4 but there was an "appreciable" reduction when the ozonation was performed at pH 9.0.

Ozonation tests were performed on the wastewater after filtration through the laboratory dual media filter. In the first series of ozonation tests, KCN was added to raise the cyanide concentration to 2 mg/l and 180 mg/l NaOH was added to raise the pH to 10. The wastewater was ozonated using the 1100 ml ozone reaction bulbs. The results of these tests are listed in Table A-1.9. The low ozone dosages used in these tests had little effect on ammonia concentration but was effective in reducing cyanide to the desired low levels. The addition of NaOH caused a precipitate to form. Ozonation caused the precipitate to change in color from white to brown. After ozonating, the pH of the samples was reduced to 7.0 with H_2SO_4 . The suspended solids concentration of the neutralized samples was about 200 mg/l; substantially greater than the suspended solids concentration (13 and 19 mg/l) measured in the dual media filter effluent in previous tests.

A second series of ozonation tests was performed on wastewater using higher ozone dosages. The ozone absorption flasks were used for these tests. Prior to ozonation, the dual media filtrate was spiked with KCN and phenol. In addition to the ozonation tests, one test (test 0-8) was performed passing only oxygen through the sample at the same flow rate and for the

TABLE A-1.9. RESULTS OF LOW DOSAGE OZONATION
OF WASTEWATER AT pH 10.0

Test Conditions:				
Wastewater filtered through dual media filter. KCN added to increase cyanide concentration to 2 mg/l. Filtrate pH raised to 10.0 with 180 mg/l NaOH. Samples were ozonated using the reaction bulb method. After ozonation, pH was reduced to 7 with H ₂ SO ₄ .				
Test No.	0-3	0-1	0-2	
Ozone dosage, mg/l	9	16	37	
pH after ozonation	9.55	9.3	9.45	
H ₂ SO ₄ used, mg/l	-	98	99	
Analyses	(Feed) ^a			
pH	6.9	7.1	7.0	7.0
Cyanide, mg/l	2.0	0.20	0.15	0.10
Ammonia, mg/l as N	19.0	17.4	18.6	17.4
TOC, mg/l	12	11	11	11
Suspended solids, mg/l	- ^b	238	-	215

- a. DMF filtrate after spiking; before addition of NaOH.
b. Suspended solids concentration of DMF effluent was less than 20 mg/l in previous tests.

same length of time used for the highest ozone dosage. The results of these tests are listed in Table A-1.10. The lowest dosage of ozone in these tests (37 mg/l) was sufficient to reduce cyanide to less than 0.25 mg/l and phenol to less than 0.05 mg/l. The highest ozone dosage tested (203 mg/l) was needed to reduce the ammonia concentration to less than 10 mg/l. Ozonation was also effective in reducing BOD, TOC and Kjeldahl nitrogen. The analysis of the sample from test 0-8 shows that stripping and/or reaction with oxygen had little or no effect on cyanide, ammonia, phenol and TOC concentrations compared to the effect of ozone. Although oxygen alone resulted in a significant decrease in BOD, a greater decrease in BOD was achieved with one tenth the amount of oxygen combined with 37 mg/l ozone (test 0-7). In a similar manner, a combination of ozone and oxygen was much more effective in removing soluble manganese than oxygen

TABLE A-1.10. RESULTS OF HIGH DOSAGE OZONATION OF
WASTEWATER AT pH 10.0

<u>Test Conditions:</u>						
Dual media filtrate spiked with KCN and phenol. Filtrate pH raised to 10.1 with 107 mg/l NaOH. Samples ozonated using the absorption flask method. After ozonation, pH was reduced to 7 with H ₂ SO ₄ .						
Test No.	0-8 ^a	0-7	0-4	0-5	0-6	
O ₃ addition rate, mg/min.	0	53.1	53.8	52.9	53.3	
Time of test, min.	30	3	6	15	30.25	
O ₃ applied, mg	0	159	322	794	1612	
O ₃ absorbed, mg	0	55	94	176	305	
Waste volume, ml	1,500	1,470	1,500	1,500	1,500	
O ₃ dosage, mg/l	0	37	62	117	203	
Initial pH	10.1	10.1	10.1	10.1	10.1	
pH after ozonation	9.8	9.8	9.7	9.5	9.0	
H ₂ SO ₄ dosage, mg/l	131	98	98	88	39	
<u>Effluent Quality:</u>		(Feed) ^b				
pH	7.0	6.9	7.0	7.0	6.9	6.9
Cyanide, mg/l	1.8	1.8	0.24	0.11	0.22	0.23
Ammonia, mg/l as N	16.4	15.0	13.4	13.4	11.4	7.6
Phenol, mg/l	2.5	2.4	0.04	0.05	0.04	0.04
TOC, mg/l	15	15	12	14	13	10
BOD, mg/l	12	6	<2	<2	<2	<2
Suspended solids, mg/l	7	210	228	234	220	250
Kjeldahl nitrogen, mg/l	-	20.7	-	-	14.6	9.2
Soluble manganese, mg/l	2.9	1.08	0.46	0.28	0.13	0.05

a. This test was performed at the same oxygen flow rate as the other tests and for the same length of time as test 0-6 but the power to the ozonator was off.

b. Analysis of dual media filter effluent after spiking with KCN and phenol before addition of NaOH.

alone. As in the previous series of tests, the addition of NaOH caused a precipitate to form which turned brown when ozonated. The suspended solids values show that the amount of suspended solids was apparently not related to ozone dosage. In addition to the brown color of the precipitate, these high dosages resulted in the formation of a red-blue color in the sample supernatant. This color faded overnight and disappeared from the samples within a few days.

Settling tests were performed on a sample of DMF effluent that had been treated with 184 mg/l ozone at pH 10.0 and neutralized to pH 7 with 74 mg/l H_2SO_4 . The results of the ozonation test and settling tests are listed in Table A-1.11. The sample had been spiked with KCN prior to ozonation. After neutralization, a settling test was performed to determine the solids settling rate without chemical treatment. No effluent samples were taken from this test.

The sample was then treated with a cationic polyelectrolyte and another settling test performed. Vigorous flocculation was needed for good floc formation. The polyelectrolyte increased the settling rate from 0.015 to 0.064 m/min (.05-.21 fpm).

A third series of ozonation tests was performed on the wastewater using spiked DMF filtrate without the addition of NaOH. These tests were run at low ozone doses using the ozone reaction bulbs. The results of these tests are listed in Table A-1.12. The minimum dosage tested (2.5 mg/l) reduced the phenol concentration to less than 1 mg/l. A dosage of 11 mg/l ozone was needed to lower the cyanide dosage to less than 0.25 mg/l. The ammonia concentration was unchanged. No precipitate was formed.

Brine Treatment

The brine from the reverse osmosis test was also treated with ozone. The brine sample was spiked with phenol and with potassium ferricyanide [$K_3Fe(CN)_6$]. The pH of the brine was adjusted to 10 to enhance ammonia removal. The brine was ozonated using the absorption flask procedure. In the first series of tests (Table A-1.13), the brine pH dropped substantially during ozonation. A second series of tests (Table A-1.14) was performed where the pH of the brine was adjusted periodically during ozonation to maintain the pH at about 10. Removal of ammonia was slightly better when the pH was maintained at 10. This effect is illustrated in Figure A-1.3 along with the pH of the brine. At an ozone dosage of 8.2 mg/l, the ammonia concentration was reduced to 13 mg/l. Phenol was reduced to less than 0.25 mg/l in both series of tests at the lowest ozone dose tested. Acceptable cyanide concentrations, however, were not achieved in any of the tests. This was probably due to the fact that the cyanide added was in the form of the ferricyanide, which is extremely difficult to destroy.

As in the case of the DMF effluent, the adjustment of the brine pH to 10 caused a precipitate to form. The precipitate turned brown when ozonated. After ozonation and neutralization, the brine had a suspended solids concentration of about 400 mg/l. The results of settling tests performed on the

TABLE A-1.11. SETTLING TEST RESULTS OF OZONATED WASTEWATER

Test Conditions:

DMF filtrate spiked with KCN. Filtrate pH raised to 10.1 with 220 mg/l NaOH. Sample ozonated using absorption flasks. After ozonation, pH reduced to 7 with 74 mg/l H_2SO_4 . Settling tests performed on neutralized sample.

	<u>Spiked DMF effluent</u>	<u>Ozonated sample</u>
Ozone dosage, mg/l	-	184
pH	-	7.0
Suspended solids, mg/l	4	242
Turbidity, FTU	18	85
Cyanide, mg/l	0.98	0.14
Ammonia, mg/l as N	15.8	10.8
TOC, mg/l	17	-

Settling Tests

Betz 1100, mg/l	0	0.5
Mix time, min.	-	2.5
Settling rate, m/min	0.015	0.064
Detention time, min	30	30
Scum volume, ml/l	Trace	Trace
Sludge volume, ml/l	Trace	2
<u>Effluent:</u>		
Suspended solids, mg/l	-	4
Turbidity, FTU	-	2.5
Ammonia, mg/l	-	10.8
TOC, mg/l	-	11

TABLE A-1.12. RESULTS OF LOW DOSAGE OZONATION
OF WASTEWATER AT pH 7.7

Test Conditions:

DMF filtrate spiked with KCN and phenol. Sample ozonated using ozone reaction bulbs. No pH adjustment before or after ozonation.

Test No.		0-10	0-9	0-11	0-12
Ozone dose, mg/l		2.5	5.5	11	22.5
Initial pH		7.7	7.7	7.7	7.7
Final pH		7.7	7.6	7.5	7.3
Analyses:	(Feed)				
Cyanide, mg/l	2.54	1.95	0.63	0.17	0.32
Ammonia, mg/l as N	18	18	18	19	19
Phenol, mg/l	1.80	0.80	0.30	<0.25	<0.25
TOC, mg/l	23	25	25	25	25
Suspended solids, mg/l	13	4	3	9	9

TABLE A-1.13. RESULTS OF OZONATION OF REVERSE OSMOSIS
BRINE AT INITIAL pH OF 10.0

Test Conditions:

Reverse osmosis brine was spiked with phenol and $K_3Fe(CN)_6$. The brine pH was raised to 10.1 with 367 mg/l NaOH. Samples were ozonated using the absorption flask method. There was no pH adjustment during ozonation. After ozonation, samples were neutralized with H_2SO_4 .

Test No.	OB-6	OB-5	OB-4	OB-1	OB-2	OB-3
O_3 Addition rate, mg/min	38.2	38.2	38.2	57.6	57.5	57.7
Time of test, min.	2	3	6	10	20	30
O_3 applied, mg	76	115	229	576	1150	1730
O_3 absorbed, mg	55	77	135	238	475	640
Waste volume, ml	960	1000	1000	1000	1000	1000
O_3 dosage, mg/l	57	77	135	238	475	640
Initial pH	10.0	10.0	10.0	10.0	10.0	10.0
pH after ozonation	9.2	9.1	9.0	8.6	7.6	7.5
H_2SO_4 dosage, mg/l	127	122	103	51	15	10
Analyses:	(Feed)					
pH	7.7	7.0	7.0	7.0	7.0	7.0
Cyanide, mg/l	3.8	0.9	4.1	4.1	3.7	1.0
Ammonia, mg/l as N	65	55	52	43	37	32
Phenol, mg/l	4.1	<0.25	<0.25	<0.25	<0.25	<0.25
TOC, mg/l	-	36	25	25	23	24
Suspended solids, mg/l	-	460	430	400	390	380

TABLE A-1.14. RESULTS OF OZONATION OF REVERSE OSMOSIS
BRINE AT MAINTAINED pH of 10

Test Conditions:

Reverse osmosis brine was spiked with phenol and $K_3Fe(CN)_6$. The pH of the brine was adjusted to 10.0 with 360 mg/l NaOH. Samples were ozonated using the absorption flask method. Ozonation was interrupted periodically to re-adjust pH to about 10. After ozonation, samples were neutralized with H_2SO_4 .

Test No.	0B-10	0B-9	0B-8	0B-7
O_3 addition rate, mg/min	50.2	50.2	50.2	53.4
Time of test, min	5	10	20	30
O_3 applied, mg	251	502	1012	1600
O_3 absorbed, mg	146	274	508	812
Sample volume, ml	1000	1000	1000	1000
O_3 dosage, mg/l	146	274	508	812
Initial pH	10.0	10.0	10.0	10.0
Total NaOH dosage, mg/l	360	400	520	720
pH after ozonation	9.9	10.0	9.1	10.1
H_2SO_4 dosage, mg/l	255	245	245	235

Analyses:	(Feed)				
pH	7.7	7.1	6.9	6.1	6.9
Cyanide, mg/l	3.8	1.25	0.31	0.57	0.55
Ammonia, mg/l as N	65	52	42	27	13
Phenol, mg/l	4.1	<0.25	<0.25	<0.25	<0.25
TOC, mg/l	23	13	7	7	5
BOD, mg/l	-	<2	<2	<2	<2
Suspended solids, mg/l	16	364	400	490	430

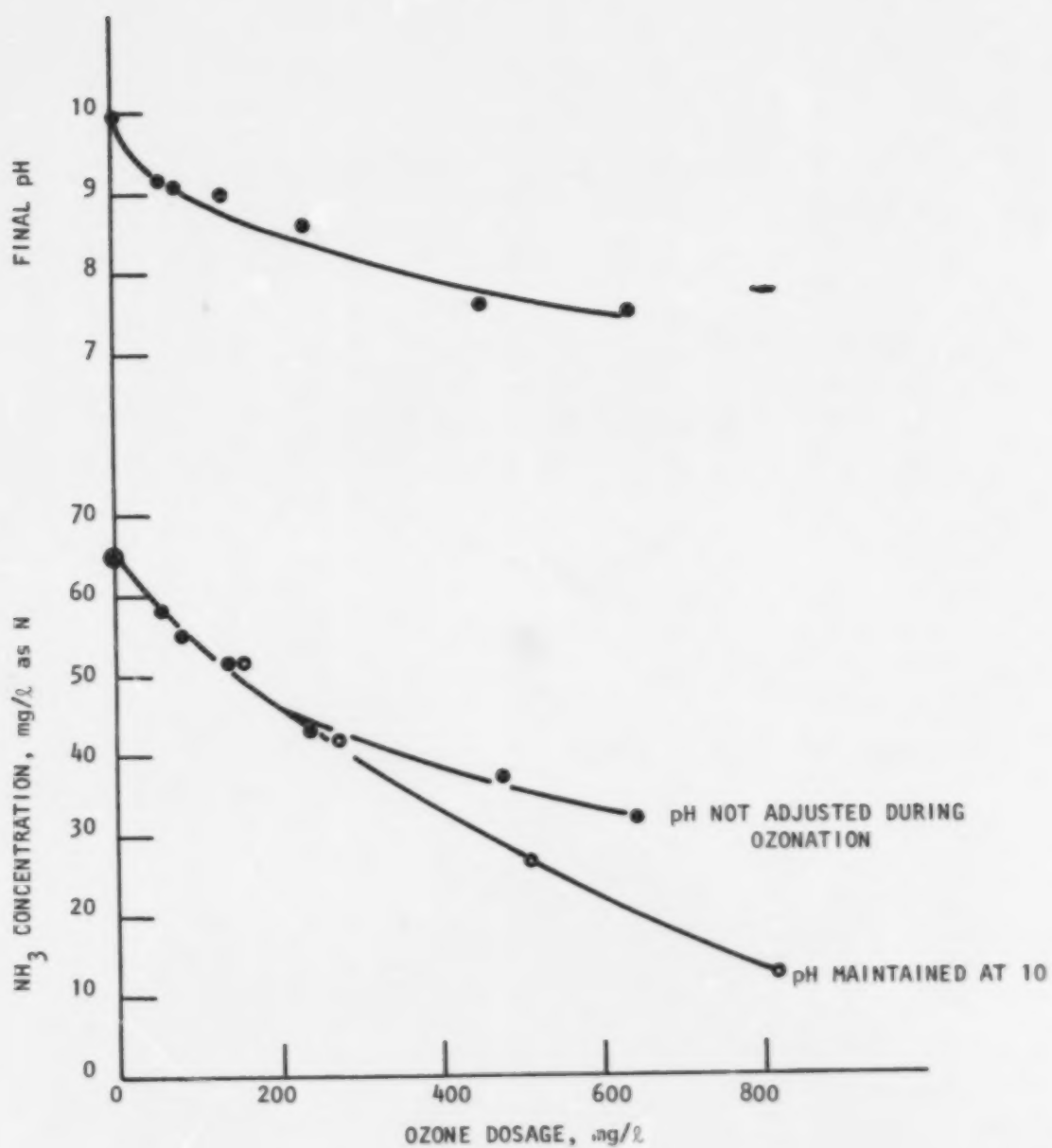


Figure A-1.3. Effect of ozone dosage on ammonia concentration and pH of reverse osmosis brine.

neutralized brine are listed in Table A-1.15. These solids had a faster settling rate and a higher sludge volume than the solids from the treatment of DMF effluent.

Recommendations for Pilot Study

Removal of phenol and free cyanides from blast furnace wastewater can be accomplished at low ozone dosages (about 11 mg/l) and without pH adjustment. Removal of ammonia, however, requires high ozone doses (about 200 mg/l) and adjustment of pH to 10.0 with NaOH at dosages of about 100 to 200 mg/l. After ozonation, the wastewater requires neutralization and settling to remove suspended solids. About 40 to 100 mg/l H_2SO_4 may be needed for neutralization. A settling rate of 0.015 m/min can be expected. Treatment with 0.5 mg/l Betz 1100 increases the settling rate to 0.064 m/min. A sludge volume of 2 ml/l and an effluent suspended solids concentration of 4 mg/l can be expected after settling of the chemically treated waste. The ozonated waste is highly colored after ozone treatment but the color disappears on standing. The presence of substantial amounts of ferricyanide in the wastewater may make it difficult to reach the required 0.25 mg/l concentration.

Ozone dosages in excess of 800 mg/l will be needed to reduce ammonia concentrations of reverse osmosis brine to the required 10 mg/l level. At these high dosages, periodic addition of NaOH during ozonation is required to keep the pH near 10. A total dose of about 720 mg/l NaOH may be needed. Neutralization of the ozonated brine requires about 250 mg/l H_2SO_4 . Under these conditions, it is expected that phenol and free cyanide can be reduced to acceptable concentrations. The presence of ferricyanides, however, may make it difficult to reach the required total cyanide concentrations. After neutralization, the suspended solids can be removed by settling. A settling rate of 0.15 m/min and a settled sludge volume of 20 ml/l can be expected. The use of 0.5 mg/l Betz 1100 is expected to increase the settling rate to 0.64 m/min and improve the effluent suspended solids concentration. The ozonated brine is highly colored, but the color fades on standing.

CHLORINATION

Results of the Chlorine Oxidation Tests and Carbon Adsorption Isotherms

The effluent concentration goals for these treatment tests were <0.2 mg/l total cyanide and <10 mg/l NH_3-N . The wastewater sampled was initially low in total cyanide so that test samples were spiked with a mixture of simple cyanide (KCN) and complex cyanide [$K_3Fe(CN)_6$].

The raw waste samples were pH adjusted with 240 mg/l NaOH to allow for efficient oxidation of cyanide to cyanate (pH 10.7 ± 0.2) and ammonia to N and CO_2 (pH 7.4 ± 0.2) (30). Chlorination breakpoint tests did not provide a clear breakpoint inflection from which to estimate efficient treatment doses (Figure A-1.4). The two chlorine treatment levels selected (Table A-1.16) were meant to bracket the theoretical reaction ratio of chlorine

TABLE A-1.15. SETTLING TEST RESULTS OF OZONATED BRINE

Test No.	1	2
Betz 1100, mg/l	0	0.5
Mix time, min	-	0.5
Settling rate, m/min	0.15	0.64
Detention time, min	30	30
Scum volume, ml/l	Trace	0
Sludge volume, ml/l	20	20
<u>Effluent:</u>		
pH	7.0	7.0
Suspended solids, mg/l	23	12

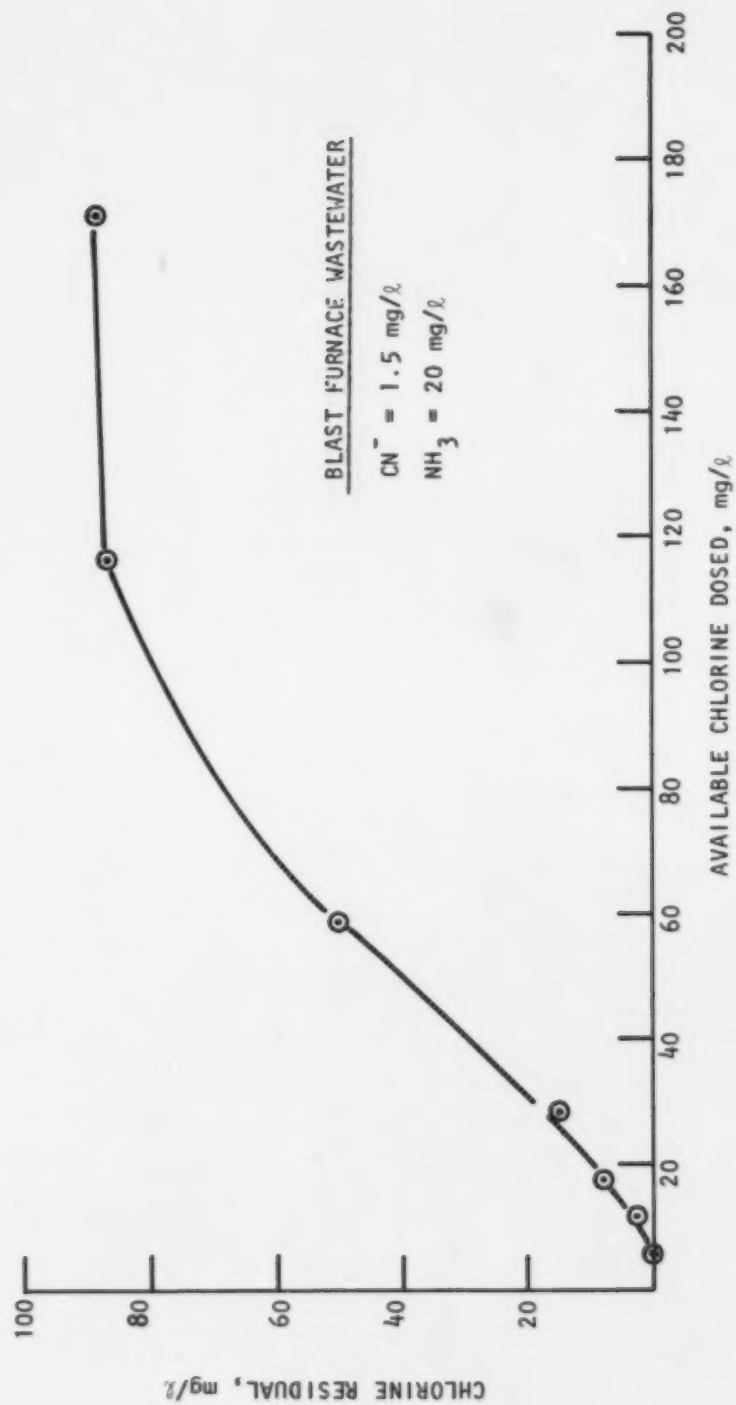


Figure A-1.4. Chlorine residual vs. chlorine dose.

TABLE A-1.16. EFFECT OF CHLORINATION ON CYANIDE AND AMMONIA REMOVAL

I. Raw Wastewater				
Test	1		2	
	Raw	Treated	Raw	Treated
Chlorine dose (available chlorine, mg/l)	0	60	0	170
pH	7.5	6.3	7.4	6.4
Suspended solids, mg/l	32	8	34	9
Total cyanide, mg/l	4.3 ^a	1.9	1.8 ^b	0.4
Ammonia-N, mg/l	17.5	16.0	19.5	6.6
Residual chlorine after sedimentation, mg/l	--	46	--	35
TOC, mg/l	16	17	16	15
II. Reverse Osmosis Brine				
Test	Raw	1	2	3
Chlorine dose, mg/l	--	280	500	670
pH	6.9	7.2	7.3	7.2
Total cyanide, mg/l	3.8 ^c	2.8	3.1	2.2
Ammonia-N, mg/l	65	29	2.1	0.3
Phenol, mg/l	4.1	<0.3	<0.3	<0.3
TOC, mg/l	27	30	30	27

- a. The raw wastewater was spiked with 2.0 mg/l CN from $K_3Fe(CN)_6$, and 1.6 mg/l CN from KCN.
- b. The raw wastewater spiked with 1.3 mg/l CN from $K_3Fe(CN)_6$.
- c. The brine was spiked with 1.9 mg/l CN from $K_3Fe(CN)_6$.

and ammonia which is 7.6 to 1 (30,31). Test No. 1 was treated with approximately 45 percent of the chlorine necessary to react with the ammonia. The results indicated that very little ammonia was removed. The total cyanide concentration of 4.3 mg/l contained 2.0 mg/l CN from the complexed form $K_3Fe(CN)_6$. The resultant cyanide concentration of 1.9 mg/l after chlorination indicates very little of the complexed cyanide was removed. Test No. 2 was chlorinated with about 1.3 times the theoretical chlorine concentration necessary to remove the initial ammonia concentration. The ammonia was reduced to an acceptable 6.6 mg/l. However, the total CN was 0.4 mg/l, or 0.2 mg/l above the acceptable concentration. The higher chlorine concentration apparently oxidized some of the 1.3 mg/l CN in the form of $K_3Fe(CN)_6$ but the reaction did not proceed far enough to meet effluent requirements. Other chlorination tests in which longer contact times were evaluated also failed to reduce complex cyanide forms to acceptable levels.

After chlorination, the pH was reduced to 7.4 with 307 mg/l HCl. The precipitate which formed during alkaline chlorination was then removed by chemical treatment and settling. Treatment with 50 mg/l ferric chloride and 0.5 mg/l Betz 1100 produced a floc which settled at a rate of 0.061 m/min (0.2 fpm). A chemical mix time of 30 seconds and a flocculation time of 2.5 minutes were used. The settled sludge volume was less than 10 ml/l after a 30 minute laboratory detention time.

A series of chlorination tests were also conducted on the brine from the RO module (Table A-1.16, Part II). The brine was spiked with complexed CN and phenol. The results were similar to the raw wastewater chlorination tests. Ammonia removal did not fall below accepted concentrations until a ratio of 7.6 to 1 (Cl_2 to NH_3) was met or exceeded. Cyanide concentration did not meet effluent guideline requirements. Chlorination did remove phenol to less than 0.3 mg/l at all concentrations investigated.

Samples from the chlorination tests were chemically treated and filtered. The waste was then treated with activated carbon to remove residual chlorine, ammonia and cyanide. The results from the initial chlorination test (Table A-1.17) indicated that cyanide and chlorine could be removed with activated carbon. Ammonia was not adsorbed.

A raw wastewater spiked with 0.6 mg/l CN from $K_3Fe(CN)_6$ which was not chlorinated was evaluated with activated carbon. The estimated saturation value was about 1/2 that found for CN after chlorination. However, the CN was removed to levels acceptable for discharge in both studies.

Recommendations for Pilot Tests

It is recommended that alkaline chlorination, settling, and carbon adsorption be investigated in the pilot test phase for the purpose of removing ammonia, cyanide and phenol.

Based on the sample tested, about 240 mg/l NaOH will be needed to raise the pH of the wastewater to the level needed for alkaline chlorination. A high chlorine dose (170 mg/l; 8.7 mg Cl_2 /mg NH_3N) was required to reduce the ammonia concentration to less than 10 mg/l. After a 7.5 minute laboratory

TABLE A-1.17. RESULTS OF THE CARBON ADSORPTION^a TESTSI. Effluent from chlorination test #1, initial chlorine dose 60 mg/l.

	Parameter concentration		Max. removal mg parameter/ g carbon	Estimated saturation value liter of influent/ kg carbon
	Initial	Final (at max. removal)		
Chlorine	35 mg/l	<1 mg/l	62	1,771
Total cyanide	0.7 mg/l	<0.1 mg/l	1.5	2,143
Ammonia-N	16 mg/l	14 mg/l	no sorption measured	--
TOC	15 mg/l	<1 mg/l	18	1,200

II. No Chlorination.

Total cyanide ^b	0.8 mg/l	<0.1	0.8	1,000
Ammonia-N	14	13	no sorption measured	

a. Carbon was regenerated bituminous grade.

b. 0.6 mg/l CN as $K_3Fe(CN)_6$ added.

reaction time, 307 mg/l HCl was needed to reduce the pH to 7.4.

The precipitate formed during alkaline chlorination can be removed by chemical treatment and settling. The use of 50 mg/l ferric chloride and 0.5 mg/l Betz 1100 produced a floc with a settling rate of 0.061 m/min (0.2 fpm). A 30 second chemical mix time and 2.5 minute flocculation time was satisfactory. Settled sludge volume was less than 10 ml/l.

The residual chlorine was removed by activated carbon. The cyanide remaining after chlorination was also reduced to the desired level.

It should be noted that the chlorination step alone was ineffective in reducing the cyanide concentration to an acceptable level. Activated carbon treatment of the unchlorinated sample removed cyanide but did not significantly remove ammonia. Only the combination of alkaline chlorination and carbon adsorption was effective in removing both cyanide and ammonia.

APPENDIX A-2
CONVENTIONAL POLLUTANT DATA
TABLE A-2.1. PILOT PLANT DATA SUMMARY

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: December, 1978

Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	RAW WASTEWATER					TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
							Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color units	TOC mg/l			
1														
2														
3														
4														
5														
6														
7														
8														
9														
10														
11														
12														
13														
14														
15														
16														
17		54												
18	7.80	40	18	45.7	0.56			19		5/10	5	50.4		
19		49												
20		62												
21		70												
22		49												
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: East Furnace Blowdown, Burns Harbor

Data Collected During: December, 1978

Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1														
2														
3														
4														
5														
6														
7														
8														
9														
10														
11														
12														
13														
14														
15														
16														
17														
18		6												
19														
20		12												
21		19												
22		17												
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: December, 1978

MAGNETIC FILTER EFFLUENT (RAW WASTEWATER FEED)														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1														
2														
3														
4														
5														
6														
7														
8														
9														
10														
11														
12														
13														
14														
15														
16														
17		31												
18														
19		34												
20		42												
21		29												
22														
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: January, 1979

RAW WASTEWATER														
Date	pH	SS mg/l	VSS mg/l	NH3 mg/l	CN-A mg/l	BOD5 mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color units	TOC mg/l	TKN mg/l	NO2/NO3 mg/l	TDS mg/l
1														
2														
3														
4		98												
5														
6		43												
7		62												
8														
9		49												
10	7.60	33	11	46	0.30		0.040	12.2	<0.2		12	53		
11														
12	7.95	60		72.1										
13	7.10	42	6	58.6	0.36		0.038	11.2	<0.2	5/10	10	64.5		
14														
15	7.10	102	27	42.9	0.25		1.10	13.5	<0.2	5	14	47.0		
16	7.20	299	29	41.4	0.35	24	0.610	11	0.3	5	60		0.11	
17		93	13	43.9	0.57		0.034	10.5	<0.2	5	18	44	0.08	
18						11								
19		110	17	46.9	0.54	19.1	0.054	11		10	44	46.8		
20	7.70	160	17	29.9	0.33		0.025	12.5		5	38	29.6		
21														
22	7.10	127	18	38	0.49		0.060	17.2	<0.1	5	20	41.3		
23	7.10	97	13	41	0.33		0.045	15.4	0.1	5	13	43.6		
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (cont inued)

Site: Blast Furnace Blowdown, Burns Harbor

Date Collected During: January, 1979

DUAL MEDIA FILTER EFFLUENT (RAW WASTEWATER FEED)

Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CH-A mg/l	BO ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TNM mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1														
2														
3														
4		10												
5														
6		12												
7		12												
8														
9		4												
10		11												
11														
12		10	5											
13		4												
14														
15		29												
16		33												
17		40												
18		19												
19		10												
20														
21														
22		2												
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blended, Burns Harbor
Data Collected During: January, 1979

MAGNETIC FILTER EFFLUENT (RAW WASTEWATER FEED)																
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CH-A mg/l	BOD ₅ mg/l	Pheno mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l		
1																
2																
3																
4																
5																
6																
7																
8																
9		27														
10		21	8													
11																
12																
13																
14																
15																
16																
17																
18																
19		32														
20		22														
21																
22		8														
23																
24																
25																
26																
27																
28																
29																
30	*															
31																

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Date Collected During: January, 1972

O ₃ EFFLUENT (FILTERED WASTEWATER FEED)														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CH-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1														
2														
3														
4														
5														
6														
7														
8														
9				34	0.02		0.204				6			
10														
11														
12														
13	7.90	1	<1	3.1	<0.01		0.010	10.8	<0.2	5/10	<2	1.1		
14														
15	7.80	3	<3	29.0	<0.01		0.066	15.3	<0.2	10	<2	31.3		
16	7.40	9	2	13.7	<0.01	5.9	<0.001	11.5		0/5	3		25.5	
17		7	3	9.4	<0.01	<5	0.960	11	<0.2	0/5	6	4.8	34.5	
18														
19	7.60	5	2	13.0	<0.01		0.002	10.5	0.3	5	4	11.0	38.5	
20	6.90	15	5	3.4	<0.01		0.005	12	<0.2	5	5	0.67		
21	7.50	7	1	32	0.50		0.660	15.6	<0.1	10	<1	35.2		
22	7.60	16	5	10	0.03		0.190	16.2	0.1	10	<1	12.7		
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: February, 1979

Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	RAW WASTEWATER									
					Ca-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TEN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1	7.70	49	9	32.8	0.12		0.025	12.5	<0.1	0/5	13	33.3	0.03	1080
2														
3														
4	7.80	55	7	38.3	0.12		0.029	11.5	<0.2	15	26	40.3	0.06	1238
5														
6	7.80	73	8	32.9	0.19		0.024	13.5	<0.2	15	35	35.3	0.06	1120
7	7.80	51	11	29.0	0.10	21	0.020	13.1	0.25	5/10	17	31.4	0.09	1100
8	7.55	85	15	29.2	0.02		0.027	11.6	0.29	20/30	33	31.5		1180
9		101	27											
10	7.30	66	18	30.4	0.01		0.015	13.4	<0.2	5/10	35	31.4	0.09	1058
11	7.70	62	23	32.4	0.02		0.017	13.7	<0.2	15	34	33.5	0.10	1001
12	7.70	75	17	35.9	<0.01		0.020	14.8	<0.2	15	28	36.5	0.10	1125
13	7.80	870	360	36.3	0.17		0.011	21.4	<0.2	15	180	39.8	0.10	1474
14	7.90	3480	630	17.0	0.18	<15	0.044	13.7	<0.2	10	1340	46.6	0.08	1433
15	7.90	3595	335	34.2	0.43	<15	0.030	16.2	<0.2	20	620	48.2	0.11	1395
16						<15								
17														
18														
19	7.30	26	6	33.3	0.34		0.019	21.4	<0.2	0/5	35	37.9	0.07	1210
20	7.80	95	18	34.0	0.23		0.026	18.9	<0.2	5	45	40.2	0.06	1160
21	7.90	101	22	38.4	0.28	10.4	0.017	16.9	<0.2	0	62	38.9	0.00	1255
22	8.00	122	22	37.8	0.23	18.2	0.020	18.0	<0.2	0/5	46	42.3	0.07	1280
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blastdown, Burns Harbor

Data Collected During: February, 1972

ALKALINE CHLORINATION CLARIFIED EFFLUENT														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1	7.20	7	6	5.2	0.02		0.386	14.4	<0.1	0/5	15	5.9	0.59	1840
2														
3														
4	7.30	8	1	1.7	<0.01		0.014	10.5	<0.2	10	22	2.0	2.30	2820
5														
6	7.30	16	4	0.48	<0.01		0.014	12.5	<0.2	20	12	1.6	1.96	2280
7	7.10	16	7	0.67	<0.01	4.0	0.011		<0.2	5/10	11	2.6	0.95	1810
8	7.00	16	6	0.67	<0.01		0.004	11.0	<0.2	5/10	10	1.2	2.20	2350
9														
10	6.30			10.3	<0.01		0.008	12.4	<0.2	10	13	11.8	0.43	2152
11	6.70	25	10	5.2	<0.01		0.009	13.4	<0.2	15	15	5.6	0.49	2142
12	6.70	11	8	7.4	0.11		0.012	14.2	<0.2	15	15	7.9	0.74	2527
13	7.20	33	13	4.8	<0.01		0.004	19.2	<0.2	15	19	3.0	1.42	3053
14	7.20	35	18	6.95	<0.01	17.1	0.013	13.6	<0.2	10	26	8.5	0.59	2689
15	6.80	33	10	10.9	<0.01	29	0.009	15.9	<0.2	15	20	13.4	0.57	2456
16						14.5								
17														
18														
19	6.40	29	11	7.35	<0.01		<0.001	20.8	<0.2	5/10	32	9.9	<0.01	2020
20	7.40	25	7	6.15	<0.01		0.005	18.0	<0.2	0	46	9.5	<0.01	2025
21	7.40	37	11	11.9	<0.01	13.8	0.005	16.8	<0.2	0	32	14.8	0.46	2155
22	7.10	36	9	4.80	<0.01	20	0.005	16.9	<0.2	0	25	6.7	0.66	2203
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowsdown, Burns Harbor

Data Collected During: February, 1979

ACTIVATED CARBON EFFLUENT														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	N ₂ /NO ₃ mg/l	TDS mg/l
1														
2														
3														
4														
5														
6	7.20	6	2	1.0	<0.01		0.216	12.5	<0.2	5	8	1.2	0.87	2260
7	7.20	9	5	0.65	<0.01	7.5	0.083	12.2	<0.2	5/10	10	2.0	0.99	2090
8	6.70	8	5	0.65	<0.01		0.360	10.0	<0.2	20/30	10	1.1	0.94	2310
9														
10	6.60	44	27	13.8	<0.01		0.545	13.6	<0.2	5/10	23	14.9	0.51	2181
11	6.80	23	17	6.7	<0.01		0.045	13.4	<0.2	12	24	7.0	0.44	2182
12	6.70	7	5	10.8	0.06		0.030	13.4	<0.2	10	12	11.1	0.72	2467
13	7.00	9	6	3.4	<0.01		0.116	18.7	<0.2	15	13	4.6	1.31	2999
14	7.45	14	9	8.15	<0.01	9.1	0.123	14.6	<0.2	5/10	14	10.4	0.74	2683
15	7.00	18	10	15.9	<0.01	10.3	0.007	15.7	<0.2	5	14	18.5	0.57	2526
16						4.0								
17														
18														
19	7.30	26	14	10.7	0.06		0.004	21.1	<0.2	5	39	11.9	0.29	2040
20	7.20	23	9	7.95	<0.01		0.002	17.9	<0.2	0	21	10.4	0.19	2070
21	7.20	21	7	14.0	<0.01	20	0.002	17.3	<0.2	0	29	16.9	0.45	2100
22	7.20	29	6	5.20	<0.01	5.5	0.005	17.2	<0.2	0/5	32	7.2	0.62	2310
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

RAW WASTEWATER															
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l	
1	7.78	573	70	38.2			0.023	9.6	<0.2	5/10	4	44.2	0.08	1372	
2	7.80	220	23	31.4		6	0.022	11.8	<0.2	0/5	8	33.3	0.11	1093	
3	7.70	37	13	37.7	0.09		0.022	12.9	<0.2	0/5	28	39.8	0.08	1400	
4	8.00	161	24	40.3	0.07		0.015	12.4	<0.2	0/5	10	42.4	0.06	1330	
5	7.95	101	16	38.2	0.06		0.018	12.7	<0.2	0/5	12	39.5	0.10	1336	
6	8.10	170	25	36.4	0.03		0.022	13.6	<0.2	0/5	6	38.9	0.09	1419	
7	7.80	147	35	40.8	0.06		0.018	8.97	<0.2		26			1318	
8	8.00	63		46.0	0.05		0.019	9.8	<0.2	0/5	12			1451	
9	8.10	84		53.4	0.24	4	0.019	6.44	<0.2		14			1490	
10	7.90	368		53			0.009	9.7	<0.2					1350	
11	8.10	231		38					<0.2		45			1190	
12	8.00	98		63			0.018	9.3	<0.2		19			1230	
13	8.10	99		36.5							6			1200	
14	7.8	61		43			0.033	9.8	<0.2		7			1266	
15	8.3	315		45	0.32		0.047	8.1	<0.2	0	40			1240	
16	8.3	91		45	0.16		0.023	7.5	<0.2		6			1208	
17															
18	8.0	31		35.5					<0.2		8			1249	
19	8.2	91		30							6	30.8		1253	
20	8.1	229	19	31.5	<0.07		0.014	9.2	<0.2	0	33	33.1		1338	
21	8.3	78	7	36	<0.01		0.013	9.6	<0.2		4	26		1187	
22	8.2	99	15	29	0.06		0.018	11.4	<0.2		12	27		1394	
23	8.3	121	14	52	0.23		0.021	12.7	<0.2		15	51		1433	
24	8.2	95	8	31.5	0.68		0.008	8.2	<0.2		20	33.6		1330	
25	8.3	84	13	33.5	0.20		0.01	7.16	<0.2		19	30.3		1202	
26	7.8	97	11	31.5	0.10		0.018	7.03	<0.2		10	29.1		894	
27	8.1	347	30	27.5	0.02		0.025	8.21	<0.2		26	24.9		1232	
28															
29															
30															
31															

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

CHEMICAL TREATMENT EFFLUENT														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TCC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1	7.48	42	8	40.2			0.032	9.6	<0.2	0/5	6	40.8	0.10	1379
2	7.35	29	5	29.6			0.017	9.4	<0.2	0/5	6	31.2	0.10	1100
3	7.78	21	10											1420
4	7.83	26	12											1384
5	7.70	24	9											1346
6	7.70	23	8											1419
7														
8		27									4			
9		20				5					4			1374
10		28												
11		21												
12	7.70	24						7.2			4			
13	7.30	21									2			1150
14														1367
15		26									5			
16		48	8					6.0			9			
17														
18	7.4	16	4											
19		26	4											
20														
21		25	9								1			
22														
23														
24	7.6	23												
25	7.4	31									16			
26	7.3	20									12			
27	7.6	25									7			
28											7			
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

DUAL MEDIA FILTER EFFLUENT (CHEMICAL PRETREATMENT)														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CH-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1	7.35	37		40.4			0.020	9.8	<0.20	0/5	10	42.3	0.10	1358
2	7.64	57	0	32.2		<6	0.012	9.7	<0.2	0/5	6	34.2	0.11	1119
3	7.60	10	5	30.7	0.05		0.015	10.9	<0.2	0/5	2	33.7	0.10	1380
4	7.70	25	6	41.8	0.05		0.013	10.3	<0.2	0	2	43.2	0.07	1396
5	7.68	12	4	30.4	0.03		0.014	10.8	<0.2	0/5	1	35.0	0.95	1331
6	7.85	6	4	48.5	0.02		0.028	8.0	<0.2	0/5	2	50.2	0.10	1418
7	7.40	15	6	43.0	0.02		0.011				2			1376
8	7.60	3	1	43.4	<0.01		0.010	8.58		0	1			1419
9	7.60	7		44.8	<0.01	<2	0.008	6.90			1		0.07	1648
10	7.50			52	<0.01		0.009	8.5		0/5	2			1400
11	7.90	19		36.5	<0.01		0.014	9.9			5			1210
12		7												
13		9												
14	6.5	13		46	0.27		0.151	7.8	<0.2		2	43.4		1353
15		10												
16	7.1	8		47	0.04									1340
17														
18	7.5	3		37										
19	6.7	6		33.5							5			1379
20	6.6	36	1	38.5	<0.01		0.174	5.5	<0.2	0	4	39.7		1305
21	6.6	12	3	4.5	<0.01		0.176	3.7	<0.2		1	3.6		1422
22	6.7	9	3	26	<0.01		0.025	5.2	<0.2		2	26.4		1320
23	6.8	8	3	27	<0.01		0.015	6.6	<0.2		1	27		1364
24	6.7	13	2	31.5	0.02		0.011	5.2	<0.2		7	32.5		1406
25		18												1322
26	7.1	21	8	30.5			0.012	4.56	<0.2		7	30.6		1157
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

REVERSE OSMOSIS PRODUCT														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Penol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1														
2	5.35	2	1	3.3	<0.01	<6	0.071	0.28	<0.2	0	1	5.5	<0.01	15
3	5.60	1	1	9.0	<0.01		0.122	0.3	<0.2	0	<1	11.2	<0.01	41
4	5.65	5	2	3.4	<0.01		0.151	0.3	<0.2	0	<1	4.5	<0.01	9
5														
6	5.95	5	4	17.3	<0.01		0.128	0.8	<0.2	0	1	18.6	<0.01	30
7	6.30			30.4	<0.01		0.143		<0.2	0	<1			157
8	6.60	3		21.4	<0.01	<2	0.030	0.66		0	<1		<0.01	134
9	5.90			25.5	<0.01									110
10														
11	6.70			21.5	<0.01						<1			180
12	5.50			19	<0.01		0.072	0.73			1			85
13														
14	6.0	6	2	28	<0.01		0.023	0.65			4	24	0.03	71
15	7.3			23.3	<0.01									112
16														
17	8.2	1		15	<0.01		0.01	0.41		0	<2		0.04	172
18				14							<2			148
19				12.3							<2			149
20														
21	5.8			16.5	<0.01		0.016	0.32	<0.2		<1	13.0		195
22	7.1			15.5	<0.01		0.007	0.61	<0.2		<1	16.5		229
23	6.3			20	<0.01		0.018	0.68	<0.2		4	13.0		106
24														
25	6.8			19.8	<0.01		<0.001	0.78	<0.2		9	19.6		141
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

REVERSE OSMOSIS BRINE														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1														
2														
3														
4														
5														
6	7.05			109.0	0.04		0.580	13.4	<0.2	0/5	13	110.2	0.31	4951
7	7.20			130.5	0.01		0.535		<0.2	5	13			5119
8	7.20	26		117.0	<0.01	7	0.563			0/5	12		0.27	4794
9	7.30			136	<0.01			13.0			13		0.55	
10														
11	7.30			100	<0.01						17			4840
12	7.10			159	<0.01		0.595	6.6			13		0.60	4540
13														
14	7.2			115	0.51		0.029	21.7	<0.02		20			5262
15	7.5			172	0.38		0.228	21.3		5	14			4665
16														
17	6.7	29		60	<0.01		0.053	22.6		0	27		0.54	5114
18				152							10			5311
19				87			0.66				10			5616
20														
21	6.5			90	<0.01		0.023	18.9	<0.2		8	81		5630
22	6.7			123	<0.01		0.022	35.2	<0.2		10	120		7655
23	6.5			89	0.08		0.011	9.4	<0.2		16	77		5500
24														
25	7.4			73	<0.01		0.016	11.44	<0.2		10	66.6		4310
26														
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

ACTIVATED CARBON EFFLUENT (ALKALINE CHLORINATED REVERSE OSMOSIS BRINE FEED)														
Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	NO ₃ /NO ₂ mg/l	TDS mg/l
1														
2														
3														
4														
5														
6														
7														
8														
9														
10														
11														
12														
13														
14														
15														
16														
17														
18														
19														
20														
21	8.1	34	9	25	<0.01		0.137	20.9	<0.2		1	24.1		7878
22				4.7	<0.01		0.02		<0.2		<1	5.9		7236
23	6.7	19		2.5	<0.01		<0.001				6	3.4		7958
24														
25				30	0.04		0.007	9.5			3	28.3		6800
26	6.6	42	6	20	<0.01		0.012	11.44			4	19.6		6205
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

Date	pH	ALKALINE CHLORINATION CLARIFIED EFFLUENT (REVERSE OSMOSIS BRINE FEED)										TKN mg/l	NO ₃ /NO ₂ mg/l ³	TDS mg/l
		SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l			
1														
2														
3														
4														
5														
6														
7														
8														
9														
10														
11														
12														
13														
14														
15														
16														
17														
18														
19														
20	6.8	88	14	1.8	<0.01		0.002	21.3	<0.2	0	7			
21	6.7	54	10	4.2	<0.01		0.01	20.1	<0.2		5			7996
22	7.0			2.4							5			7424
23	5.8	42	5	<0.05	<0.01		0.015	21.9	<0.2		7	1.1		9482
24	6.9	36	9	20.3	<0.01		0.018	8.76			4	19.6		8023
25				23.8	0.03		0.005	10.5			13			7061
26	6.6	34	5	16.8	<0.01		0.005	10.56			15	16.0		6105
27														
28														
29														
30														
31														

TABLE A-2.1. (continued)

Site: Blast Furnace Blowdown, Burns Harbor

Data Collected During: March, 1979

OZONE CLARIFIED EFFLUENT (REVERSE OSMOSIS BRINE FEED)

Date	pH	SS mg/l	VSS mg/l	NH ₃ mg/l	CN-A mg/l	BOD ₅ mg/l	Phenol mg/l	Fluoride mg/l	Sulfide mg/l	Color Units	TOC mg/l	TKN mg/l	NO ₂ /NO ₃ mg/l	TDS mg/l
1														
2														
3														
4														
5														
6														
7	6.90	38		39.0	<0.01		<0.01	17.4		0	22			9795
8	7.80	31		91.5	<0.01	9	0.002	19.3			7		32.0	8669
9	7.8	37		75	<0.01		0.018	16.3					48	8073
10														
11	6.60	39		57	<0.01			22		0/5	13			7750
12														
13	8.40	74		69	<0.01		0.012	18.7		0/5	7		33	4740
14	6.3	78		42	<0.01		0.007	27.6			h			10626
15	9.9	53		42	<0.01		0.023	9.1		0	6		85	8353
16	8.5	65		47	<0.01		0.056	14		0	4			9652
17														
18														
19														
20														
21														
22														
23														
24														
25														
26														
27														
28														
29														
30														
31														

TABLE A-2.2 BLAST FURNACE BAT EFFLUENT LIMITATIONS

Date	Plant discharge		TSS		Cyanide		Pheno		Ammonia		Sulfide		Fluoride	
	Product	Flow	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day
12/13/78	13,479	0.989	42.49	350.45	0.424	3.5	0.050	7.01	17.00	140.18	0.523	4.313	34.00	280.36
12/14/78	12,698	1.004	36.52	310.15	0.365	3.30	0.730	6.60	14.61	132.06	0.450	4.063	29.22	264.12
12/15/78	13,341	1.123	37.04	346.87	0.370	3.47	0.741	6.94	14.81	138.75	0.456	4.269	29.63	277.49
12/16/78	13,016	1.278	31.75	338.42	0.318	3.38	0.435	6.77	12.70	135.37	0.391	4.165	25.40	270.73
12/17/78	12,953	1.007	40.10	336.77	0.401	3.37	0.802	6.74	16.04	134.71	0.494	4.145	32.08	269.42
12/18/78	12,498	0.827	47.11	324.95	0.471	3.25	0.942	6.50	18.84	129.98	0.580	3.999	37.69	259.96
12/19/78	13,100	1.177	34.70	340.60	0.347	3.41	0.694	6.81	13.83	136.24	0.427	4.192	27.76	272.48
12/20/78	12,700	1.147	34.52	330.20	0.345	3.30	0.690	6.60	13.81	132.03	0.425	4.064	27.61	264.16
12/21/78	12,564	1.119	35.00	326.67	0.350	3.27	0.700	6.53	14.00	130.67	0.431	4.020	28.00	261.33
1/3/79	10,764	1.548	21.63	279.06	0.217	2.50	0.434	5.60	8.67	111.95	0.267	3.444	17.34	273.89
1/4/79	11,498	1.609	22.28	298.95	0.223	2.99	0.446	5.98	8.91	119.58	0.274	3.680	17.82	239.16
1/5/79	11,178	1.499	23.25	290.63	0.233	2.91	0.465	5.31	9.30	116.25	0.286	3.577	18.60	232.50
1/6/79	12,459	1.475	26.33	323.93	0.263	3.24	0.527	6.48	10.53	129.57	0.324	3.987	21.07	259.15
1/7/79	11,689	1.357	26.85	303.91	0.269	3.04	0.537	6.08	10.74	121.57	0.331	3.740	21.48	243.13
1/8/79	10,505	1.318	24.85	273.13	0.248	2.73	0.497	5.46	9.94	109.25	0.306	3.362	19.88	218.50
1/9/79	12,812	1.381	28.92	333.11	0.289	3.33	0.570	6.66	11.57	133.24	0.356	4.100	23.14	266.49
1/10/79	12,044	1.396	26.90	313.14	0.269	3.13	0.530	6.26	10.76	125.26	0.331	3.854	21.52	250.52
1/11/79	11,904	1.422	26.10	309.50	0.261	3.10	0.522	6.19	10.44	123.80	0.321	3.810	20.88	247.60
1/12/79	11,017	1.406	24.42	286.44	0.244	2.86	0.489	5.73	9.77	114.58	0.301	3.53	19.54	229.15
1/13/79	10,685	1.587	20.99	277.81	0.210	2.78	0.420	5.56	8.40	111.12	0.258	3.42	16.79	222.25
1/14/79	10,916	1.587	21.44	283.82	0.215	2.84	0.429	5.68	8.53	113.53	0.264	3.49	17.15	227.05
1/15/79	9,995	1.587	19.63	259.37	0.196	2.60	0.393	5.20	7.85	103.95	0.242	3.20	15.71	207.90
1/16/79	9,873	1.470	20.94	256.70	0.210	2.57	0.418	5.13	8.38	102.68	0.258	3.16	16.75	205.36
1/17/79	11,923	1.320	28.16	310.00	0.282	3.10	0.563	6.20	11.26	124.00	0.347	3.82	22.53	248.00
1/18/79	9,795	1.350	22.61	267.37	0.226	2.55	0.452	5.09	9.04	107.15	0.278	3.13	18.10	203.74
1/19/79	10,169	1.420	27.32	264.39	0.223	2.64	0.447	5.29	8.93	105.76	0.274	3.25	17.86	211.52
1/20/79	11,664	1.520	23.92	303.26	0.233	3.03	0.479	6.07	9.57	121.31	0.294	3.73	19.14	242.61
1/21/79	12,581	1.35	29.05	327.11	0.290	3.27	0.581	6.54	11.62	130.84	0.353	4.03	23.24	261.68
1/22/79	13,180	1.40	29.35	342.63	0.294	3.43	0.587	6.85	11.74	137.07	0.361	4.22	23.49	274.14
1/23/79	12,400	1.38	23.01	322.4	0.280	3.22	0.560	6.45	11.20	128.96	0.350	3.97	22.41	257.92
1/24/79	11,660	1.25	29.08	303.16	0.291	3.03	0.581	6.06	11.63	121.26	0.358	3.73	23.26	242.53
1/25/79	12,330	1.59	24.17	320.58	0.242	3.21	0.483	6.41	9.67	128.23	0.298	3.95	19.34	256.46
1/26/79	12,834	1.48	27.03	333.68	0.271	3.34	0.540	6.67	10.81	133.47	0.333	4.11	21.63	266.95
1/27/79	9,845	1.39	22.12	256.49	0.221	2.56	0.443	5.13	8.85	102.40	0.273	3.16	17.70	205.19

(cont. next)

(cont Insured)

TABLE A-2.2. (continued)

Date	Plant discharge Product flow MGD	TSS		Cyanide		Phenol		Ammonia		Sulfide		Fluoride		
		mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	
1/23/79	13,728	1.40	30.57	356.93	0.306	3.57	0.612	7.18	12.23	182.77	0.376	4.39	24.46	285.54
1/24/79	9,955	1.439	21.56	258.83	0.216	2.59	0.432	5.18	8.63	103.53	0.266	3.19	17.25	207.06
1/25/79	12,628	1.296	30.37	328.23	0.303	3.28	0.608	6.57	12.15	131.33	0.374	4.04	24.30	282.66
1/26/79	11,984	1.35	27.67	311.58	0.277	3.12	0.553	6.23	11.07	124.63	0.340	3.83	22.14	249.27
1/27/79	12,870	1.376	29.16	334.62	0.292	3.35	0.583	6.69	11.66	133.05	0.360	4.12	23.33	267.70
1/28/79	10,773	1.405	23.90	280.10	0.239	2.80	0.478	5.60	9.56	112.04	0.294	3.45	19.12	224.08
1/29/79	12,286	1.406	27.24	319.44	0.272	3.19	0.545	6.39	10.90	127.77	0.335	3.93	21.79	255.55
1/30/79	12,112	1.298	29.09	314.91	0.291	3.15	0.582	6.30	11.64	125.96	0.358	3.88	23.27	251.93
1/31/79	11,888	1.384	26.78	309.09	0.268	3.09	0.535	6.18	10.71	123.63	0.329	3.80	21.42	247.27
2/1/79	12,244	1.503	24.11	318.34	0.241	3.18	0.482	6.37	9.64	127.34	0.297	3.92	19.29	254.68
2/2/79	10,655	1.432	23.20	277.05	0.232	2.77	0.464	5.54	9.28	110.81	0.286	3.41	18.56	221.62
2/3/79	11,350	1.435	24.66	295.10	0.246	2.95	0.493	5.90	9.86	118.04	0.303	3.63	19.73	236.08
2/4/79	11,257	1.435	24.46	292.68	0.245	2.93	0.489	5.85	9.78	117.07	0.301	3.60	19.56	234.15
2/5/79	12,120	1.61	23.47	315.12	0.235	3.15	0.469	6.30	9.38	126.05	0.289	3.88	18.78	252.10
2/6/79	11,857	1.55	23.85	308.28	0.230	3.08	0.477	6.17	9.54	123.31	0.293	3.79	19.03	246.63
2/7/79	12,349	1.22	31.56	321.07	0.315	3.21	0.631	6.42	12.62	128.43	0.308	3.95	25.24	256.86
2/8/79	10,996	1.31	26.19	285.90	0.262	2.86	0.524	5.72	10.47	114.36	0.322	3.52	20.93	228.72
2/9/79	12,287	1.35	28.37	319.46	0.283	3.19	0.568	6.39	11.35	127.78	0.349	3.93	22.70	255.57
2/10/79	10,004	1.368	22.30	260.10	0.228	2.60	0.456	5.20	9.12	104.04	0.280	3.20	18.24	208.08
2/11/79	12,509	1.249	31.22	325.23	0.312	3.25	0.624	6.50	12.49	130.09	0.384	4.00	24.58	260.19
2/12/79	11,234	1.549	22.61	292.08	0.226	2.92	0.452	5.84	9.04	116.03	0.270	3.59	18.09	233.67
2/13/79	11,688	1.559	23.45	303.89	0.234	3.04	0.468	6.08	9.38	121.55	0.288	3.74	18.70	243.11
2/14/79	12,212	1.380	27.59	317.51	0.276	3.18	0.552	6.35	11.03	127.00	0.340	3.91	22.07	254.01
2/15/79	12,480	1.411	27.57	324.48	0.275	3.24	0.552	6.49	11.03	129.79	0.339	3.99	22.06	259.58
2/16/79	11,370	1.317	26.91	295.62	0.269	2.96	0.538	5.91	10.77	118.25	0.331	3.64	21.53	236.50
2/17/79	11,957	1.074	34.71	310.88	0.347	3.11	0.694	6.22	13.86	124.35	0.428	3.83	27.77	248.71
2/18/79	10,451	1.323	24.63	271.73	0.247	2.72	0.492	5.43	9.85	108.69	0.303	3.34	19.70	217.38
2/19/79	13,338	1.284	32.38	346.79	0.334	3.47	0.648	6.94	12.95	133.71	0.399	4.27	25.91	277.43
2/20/79	12,075	1.225	30.73	313.95	0.307	3.14	0.615	6.28	12.29	125.58	0.378	3.86	24.58	251.16
2/21/79	14,309	1.326	28.94	320.03	0.289	3.20	0.579	6.40	11.58	128.01	0.356	3.94	23.15	256.03
2/22/79	12,914	1.568	25.68	335.76	0.257	3.36	0.514	6.72	10.27	134.31	0.316	4.13	20.54	268.61
2/23/79	12,522	1.803	21.65	325.57	0.217	3.26	0.433	6.51	8.66	130.23	0.267	4.01	17.32	260.46

(continued)

(continued)

APPENDIX A-3.1
TRW PRIORITY POLLUTANT FINDINGS
FINAL REPORT

PRIORITY POLLUTANT VERIFICATION ANALYSIS FOR
EPA'S STEEL PLANT MOBILE WASTEWATER
TREATMENT SYSTEM

October 26, 1979

Prepared by: R. G. Beimer
M. K. O'Rell

TRW
DEFENSE AND SPACE SYSTEMS GROUP

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1.0 SUMMARY

Priority pollutant analyses were performed on water samples from EPA's mobile wastewater treatment system which was operated on blast furnace wastewater at Bethlehem Steel's Burns Harbor plant. Samples from three different processes, namely a) ozonation, b) alkaline chlorination/activated carbon and c) reverse osmosis, were analyzed.

All three treatment systems are effective in removal of priority pollutant metals as assessed by metal concentration levels in the effluents. The three treatment processes also were effective in the reduction of total cyanides. Ozonation appears to be the most effective process based on cyanide concentration in the final effluent. All treatment systems produced effluents with cyanide concentrations of $<20\mu\text{g/l}$, the verification limit.

A recently published quality control procedure was used for verification analysis of the volatile and non-volatile organic priority pollutants employing GC/MS techniques. No volatile organic compounds were found at levels above the verification concentration limit in any of the samples. Consequently, no assessment can be made for reduction of volatile priority pollutions by these treatment processes. The non-volatile organic priority pollutants found were bis(2-ethylhexyl) phthalate and di-n-butyl phthalate. Neither compound is suspected of being formed by the operation of the blast furnace. Di-n-butyl phthalate was present at concentration levels below the verification limit.

2.0 INTRODUCTION

This document is the final report on priority pollutant analysis for EPA's mobile wastewater treatment system which was operated on blast furnace wastewater at Bethlehem Steel's Burns Harbor plant. The work was conducted by TRW for The Environmental Protection Agency, Metallurgical Processes Branch, IERL/RTP, under the direction of Task Officer R.V. Hendriks (Task Directive 105 of contract number EPA 68-02-2689, L.D. Johnson project officer).

Priority pollutant analyses (excluding asbestos and pesticides) were performed on samples from three different wastewater treatment processes. The processes included 1) ozonation, 2) alkaline chlorination followed by activated carbon and 3) reverse osmosis. Sampling was performed by personnel from the Environmental Research Center of Rexnord, Inc., Milwaukee, Wisconsin. Rexnord was also responsible for operation of the mobile treatment system. The three processes were sampled on 20 January 1979, 15 February 1979, and 14 March 1979, respectively. An inventory of the samples collected and pertinent shipping and receiving information are given in Tables 2-1 thru 2-3. During the second and third sampling efforts, some of the samples were renumbered upon receipt at TRW to avoid duplication of sample numbers. To ensure that no confusion or mislabeling occurred during renumbering, Rexnord sample numbers were maintained and are also shown in Tables 2-1 thru 2-3.

Detailed sampling information has already been provided to EPA by Rexnord, Inc. Copies of sampling information supplied to TRW by Rexnord are contained in the Appendix. Sampling and sample preservation was performed according to EPA protocol outlined in the EPA document entitled "Sampling and Analysis Procedures for Screening Industrial Effluents for Priority Pollutants" published by The Environmental Monitoring and Support Laboratory of the US EPA, as revised April 1977.

A one day delay occurred during sample shipment from the first and second sampling efforts. No impact on sample integrity was expected because the extractions of these samples were initiated immediately upon receipt. In addition, the sample temperature was maintained at or near 4°C during shipment. The NVO composite sample of raw influent collected

from the third treatment process was broken in transit and was not analyzed.

Table 2-1 Samples From Ozonation Treatment

Sample Site Description	Trial Sample Number	Rexnord Sample Number(s)	Sample Type	Fractions Sampled (Code)	Date Sampled	Date Shipped	Date Received	Comments
Influent to Treatment	8011	39179-81	Composite Composite Grab Grab	Non-Volatiles (W) Metals (N) Cyanide (B) Volatiles (V)	20 Jan 79	20 Jan 79	22 Jan 79	Triplicate Samples
		39170-2						
		39162-4						
		39147-55						
Effluent from Treatment	8012	39193	Composite Composite Grab Grab	Non-Volatiles (W) Metals (N) Cyanide (B) Volatiles (V)	20 Jan 79	20 Jan 79	22 Jan 79	
		39192						
		39191						
		39183-8						
Carriage Water	8013	39161	Grab Grab Grab Grab	Non-Volatiles (W) Metals (N) Cyanide (B) Volatiles (V)	20 Jan 79	20 Jan 79	22 Jan 79	
		39159						
		39160						
		39156-9						
Field Blank	8014	39190	NA NA NA NA	Non-Volatiles (W) Metals (N) Cyanide (B) Volatiles (V)	20 Jan 79	20 Jan 79	22 Jan 79	
		39166						
		39165						
		39189						

Table 2-2 Samples From Alkaline Chlorination/
Activated Carbon Treatment

Sample Site Description	TW Sample Number	Rexnord Sample Number(s)	Sample Type	Fractions Sampled (Code)	Date Sampled	Date Shipped	Date Received	Comments
Influent to Chlorinator	8072	532 526 529 504-509	Composite Composite Composite Grab	Non-Volatiles(W) Metals(H) Cyanide (B) Volatiles (V)	15 Feb 79	16 Feb 79	17 Feb 79	
Effluent from Chlorinator	8052	533 527 530 510-1:514-5 512-3:516-7	Composite Composite Composite Grab Grab	Non-Volatiles(W) Metals (H) Cyanide (B) Volatiles (V) Volatiles (T)	15 Feb 79	16 Feb 79	17 Feb 79	T Volatiles Stabilized with $\text{Na}_2\text{S}_2\text{O}_3$
Effluent from Carbon Treatment	8062	534 528 531 518-9:522-3 520-1:524-5	Composite Composite Composite Grab Grab	Non-Volatiles(W) Metals (H) Cyanide (B) Volatiles (V) Volatiles (T)	15 Feb 79	16 Feb 79	17 Feb 79	T Volatiles Stabilized with $\text{Na}_2\text{S}_2\text{O}_3$
Field Blank	8041	500-1 502-3	NA NA	Volatiles (V) Volatiles (T) ($\text{Na}_2\text{S}_2\text{O}_3$ added)	15 Feb 79	16 Feb 79	17 Feb 79	

Table 2-3 Samples From Reverse
Osmosis Treatment

Sample Site Description	TRM Sample Number	Record Sample Number(s)	Sample Type	Fractions Sampled (Code)	Date Sampled	Date Shipped	Date Received	Comments
Influent to Treatment	8073	1527 1528 1529 1509-14	Composite Composite Composite Grab	Non-Volatiles (W) Metals (N) Cyanide (B) Volatiles (V)	14 Mar 79	14 Mar 79	15 Mar 79	Non-Volatile Sample Broken in Transit
Reverse Osmosis Effluent	8093	1536 1537 1538 1515-7; 1521-3	Composite Composite Composite Grab	Non-Volatiles (W) Metals (N) Cyanide (B) Volatiles (V)	14 Mar 79	14 Mar 79	15 Mar 79	
Reverse Osmosis Brine	8083	1548 1549 1550 1518-20; 1524-6	Composite Composite Composite Grab	Non-Volatiles (W) Metals (N) Cyanides (B) Volatiles (V)	14 Mar 79	14 Mar 79	15 Mar 79	
Carriage Water	8003	1539 1541 1540 1542-7	Grab Grab Grab Grab	Non-Volatiles (W) Metals (N) Cyanide (B) Volatiles (V)	14 Mar 79	14 Mar 79	15 Mar 79	
Field Blank	8043	1500 1501 1502-7	NA	Metals (N) Cyanide (B) Volatiles (V)	13 Mar 79	14 Mar 79	15 Mar 79	

3.0 SCREENING ANALYSES

Limited priority pollutant screening information was available on the raw influent to treatment for this plant. Consequently, screening was performed on the influent from the first sampling effort to determine which of the priority pollutants were present at concentration levels requiring verification. The results of the screening activities are discussed in the following sections.

3.1 METALS SCREENING

Metals screening was performed using procedures outlined in the EPA document entitled "Sampling and Analysis Procedures for Screening Industrial Effluents for Priority Pollutants". The influent contained a considerable amount of suspended solids which were not fully dissolved during the recommended digestion procedure. Consequently, the samples were filtered after the digestion step and prior to analysis to remove any remaining solids. The results of the metals screening are presented in Table 3-1. The concentration values for the elements presented in Table 3-1 were obtained using the graphite furnace and standard addition techniques, with the exception of mercury which was analyzed using the cold vapor technique. The percent recoveries reported in Table 3-1 were obtained by spiking the sample prior to digestion with known amounts of the elements. The poor recovery observed for lead is unusual and no explanation can be presented for this result. Zinc was mistakenly spiked at a high concentration. No corrections for recoveries were made in this report based on recovery data. Based on the screening results, metals verification analyses were performed for Pb, Se, As, Cu and Hg on the first set of samples (see Section 4.1).

3.2 ORGANIC SCREENING

The organic screening can be divided into three parts. The first is the determination of volatile organics using the purge and trap technique described in the referenced screening procedures manual. The second is the determination of non-volatile base/neutral components which are obtained by extraction at pH 12. Third, screening for the acid fraction compounds

Table 3-1
Influent Metals Screening Results

Element	Sample Number 8011N $\mu\text{g}/\ell$	Spike Level $\mu\text{g}/\ell$	Percent Recovery	Verification Level ($\mu\text{g}/\ell$)
Pb	189	100	41	25
Se	10			10
Be	0.9;5			50
Ag	<1;<1			5
Tl	16;21			50
Zn	580	2000	130	1000
Sb	29;19			100
As	144;160			25
Ni	16			500
Cd	0.7			5
Cu	19			20
Cr	<10			25
Hg	1.5	2	80	1

is accomplished by analyzing the organic extract obtained from the aqueous sample after base/neutral extraction and pH adjustment to 2.

3.2.1 Volatile Organic Screening

Volatile organic analysis was performed by the purge and trap technique of Bellar and Lichtenburg. Analysis of the raw influent failed to show the presence of volatile organics at a concentration level requiring verification (i.e., $10 \mu\text{g}/\ell$). A trace of methylene chloride was found but its concentration was well below $1 \mu\text{g}/\ell$. Instrumental response (total ion count) was satisfactory for the two internal standards (bromochloromethane and 1,4-dichlorobutane) thereby eliminating instrumental problems as a potential source of a false negative result. To insure that these results were correct and that no VOA's were present, two actions were taken. First, VOA samples from the other collection points were analyzed. Again, only trace amounts of methylene chloride were found. Second,

instrument sensitivity to VOA priority pollutants was confirmed by analyzing a synthetic VOA sample. Satisfactory results were obtained for the mixture. It was therefore concluded that no VOA's were present at levels requiring verification.

3.2.2 Screening of Base/Neutral Fractions

Limited screening information was available from previous work for the base/neutral fraction and as a result, no screening was performed on this program. Two priority pollutants, bis-(2-ethylhexyl) phthalate and di-n-butyl phthalate, were reported as being present. Consequently, the base/neutral samples were spiked with these two phthalates prior to extraction. Subsequent verification analysis confirmed that these were the only base/neutral priority pollutants.

3.2.3 Screening of Acid Fractions

The acid fraction generally contains the priority pollutant phenols. No phenols were found in the raw influent, therefore, no verification analyses were performed on the acid fractions for the first set of samples. Samples from the second and third treatments were analyzed to determine surrogate spike recoveries. During these analyses no other acid priority pollutants were detected.

4.0 VERIFICATION ANALYSIS

This section describes the results of the priority pollutant verification analysis performed on the Burns Harbor samples. Verification analysis was performed for metals, non-volatile organics and cyanide. No volatile organics were found at levels requiring verification in any of the samples.

4.1 METALS ANALYSIS

Metals verification analyses for the elements found at concentrations above the level of the selection criteria (see Table 3-1) in the screening activity initially were performed on the samples from the first treatment process. The analyses were done using atomic absorption spectrophotometry (AAS) with the exception of mercury which was analyzed using the cold vapor technique.

Verification analyses results presented in Table 4-1 confirmed the screening results with the exception of mercury. The verification sample contained mercury at a concentration level below that required for verification. This result was confirmed during the analysis of the second treatment samples. Consequently, mercury analysis was not performed on the last set of samples. Zinc concentration levels were found to be below the level required for verification with the exception of the carriage water from the first treatment. No explanation is evident for the unexpectedly high zinc concentration in the carriage water.

The raw influent sample contained a significant amount of suspended solids. As was mentioned in the screening section, a residue remained after the suggested $\text{HNO}_3\text{-H}_2\text{O}_2$ digestion step which was removed by filtration prior to analysis. Emission spectrographic analysis performed on the residue showed that it contained additional amounts of the following priority pollutant metals: Pb, Cr, Cu, Ag and Ni. Other elements found included Ca, Al, Na, Fe, Si, and Mg with Na and Fe composing the largest fraction of material in the residue.

A second digestion procedure also was employed on the influent from the third treatment process in an attempt to dissolve all of the suspended

Table 4-1 Metals Verification Results

Treatment/Sample Description	TRW Sample Number	Rexnord Environmental Research Sample Number	As	Cr	Cu	Pb	Se	Zn	Hg
A. O ₃ Treatment									
Influent to Treatment	8011N	39170	153;173 ^a	<10 ^b	40;40;16	189	21;11	580	0.4;0.4
Effluent from Treatment	8012N	39192	<10 ^b	e	<10;14	<10	<10	800	e
Carriage Water	8013N	39159	<10	e	<10	<10	10	1000	e
B. Alkaline Chlorination/ Activated Carbon Treatment									
Influent to Treatment	8072N	526	277	e	16	1140	7	860	<0.5
Effluent from Chlorinator	8052N	527	<10 ^b	e	43	14	24	320	<0.5
Effluent from Carbon Treatment	8062N	528	<10	e	<10;10	<10	29	470	<0.5
C. Reverse Osmosis Treatment									
Influent to Treatment	8073N	1528	179	21 ^c ; (65) ^d	24 ^c ; (35) ^d	161 ^c (114) ^d	<5	190 ^c (330) ^d	e
Brine Sample	8083N	1549	11	4	1	2	<5	60	e
Effluent from Treatment	8093N	1537	<10 ^b	6	2	NO ^f	<5	<50	e
Carriage Water	8003N	1541	<10	2	4	NO ^f	<5	<50	e

a. Duplicate Analysis.

b. Concentration level required for verification is 25 µg/l.

c. Results for recommended digestion procedure.

d. Results for HF/HNO₃/HCl digestion of residue and filtrate.

e. Not analyzed, below verification limit.

f. Not detected.

solids. This procedure used HF and HNO_3 rather than $\text{HNO}_3/\text{H}_2\text{O}_2$. The HF/ HNO_3 digestion procedure also left an insoluble residue but it did result in higher concentration values for the metals with the exception of As and Pb (see Table 4-1). The lower Pb value is probably related to obtaining a non-representative sample for analysis because of the suspended solids. No As was detected in the residue by emission spectrographic analysis, therefore, its concentration should not have increased with the new digestion procedure.

All three treatment processes appear to be effective in reducing the concentrations of the priority pollutant metals. In fact, with the exception of Se in the alkaline chlorination/activated carbon treatment, all metals are below the levels required for verification in the effluents. The concentration of selenium appears to have increased slightly in the chlorination process.

Spike recovery data for the metals analysis are presented in Table 4-2. Reasonable recoveries were obtained except for zinc in sample 8093. Lead recovery was not obtained for sample 8072 because the spike level was too low ($50 \mu\text{g}/\ell$) compared to the sample concentration ($1140 \mu\text{g}/\ell$).

4.2 ORGANIC ANALYSIS

The organic analysis portion of this task was performed using the recently published document entitled "Verification Appendix to Screening Protocol", dated January 1979, prepared by the Effluent Guidelines Division. A formal copy of this document was not received prior to receipt of samples from the first treatment process, therefore a slightly modified procedure was used. In particular, surrogate spikes and continuous liquid/liquid extractors were not used. The new verification protocol was used for samples from the second and third treatment processes.

4.2.1 Volatile Organic Verification Analysis

Screening activities showed that the samples from the ozonation treatment contained no volatile organic species above the concentration level required for verification. In fact, the VOA's detected were at concentrations below $1 \mu\text{g}/\ell$. Samples from the other two treatment processes

Table 4-2
Metals Spike Recoveries

Metal	Sample Number/Description			
	8072N (Influent)		8093N (Effluent)	
	Spike Level $\mu\text{g}/\ell$	% Recovery	Spike Level $\mu\text{g}/\ell$	% Recovery
Pb	50	(1)	50	136
Se	10	140	20	97
Zn	200	130	500	46
Cu	40	140	40	85
Cr	ND ⁽²⁾	ND	50	72
As	100	117	50	86

(1) Spike level was too low in relation to sample concentration to determine percent recovery.

(2) Not determined.

were also analyzed using the purge and trap technique described in the screening manual. No volatile organic compounds were found at concentrations $>1\mu\text{g}/\text{l}$. Therefore, no VOA verification activities were performed.

4.2.2 Base/Neutral Organic Verification Analysis

Base/neutral analysis samples were prepared by extracting the aqueous samples at pH 12 with methylene chloride. For the first set of samples (ozonation treatment), the samples were extracted using separatory funnels. Subsequent samples from the other two treatments were extracted for 20 to 24 hours in a 4-l liquid/liquid continuous extractor. A 2-l aliquot of sample was diluted with an equal volume of organic free water prior to extraction in order to make the extractor function properly. The organic extracts were dried and concentrated to 2 ml prior to analysis. Spiking was performed prior to extraction by the addition of an acetone solution of the appropriate priority pollutant or surrogate spike compound. Decafluorobiphenyl was used as the surrogate spike.

For the first set of samples, no surrogate spikes were used because the new procedures were not received until after the samples were extracted. Four 2-l aliquots each of influent and effluent were analyzed as follows. Two of the four aliquots were spiked with bis(2-ethylhexyl) phthalate and di-n-butyl phthalate prior to extraction and two aliquots were extracted without being spiked. The identical aliquots are reported as duplicates in Table 4-3.

Second and third treatment samples were treated as day one samples in the verification protocol. Accordingly, three 2-l aliquots of influent and effluent were used. All of the samples were spiked with a surrogate spike, decafluorobiphenyl ($100\mu\text{g}/\text{l}$), and one of the three samples also was spiked with priority pollutants. The identical aliquots without priority pollutant spikes are reported as duplicates. Finally, the protocol also requires that duplicate injections be made on certain of the aliquots which are also identified in Table 4-3.

Concentration levels of di-n-butyl phthalate were determined to be below the level required for verification. Conversely, bis(2-ethylhexyl) phthalate concentration levels greatly exceeded the level required for

Table 4-3 NVO Analyses Results

Treatment/Sample Description	Sample Number	Di-n-butyl Phthalate			Bis-(2-ethylhexyl) Phthalate			Surrogate Spike Recovery, Percent
		Amount Found, µg/l	Spike Level, µg/l	Percent Recovery	Amount Found, µg/l	Spike Level, µg/l	Percent Recovery	
A. O ₃ Treatment								
Influent to Treatment	8011	3	30	106	288	1220	78	NA ^d
Duplicate 8011		3;4 ^a	30	79	224;240 ^a	1220	72	NA
Effluent from Treatment	8012	8	30	74	562	1220	67	NA
Duplicate 8012		8;11 ^a	30	87	435;395 ^a	1220	87	NA
Carriage Water	8013	ND ^b	30	100	80	1220	114	NA
Duplicate 8013		2;3 ^a	30	67	110;115 ^a	1220	120	NA
B. Alkaline Chlorinator/ Activated Carbon Treatment								
Influent to Treatment	8072	1	-	-	81	-	-	-
Duplicate 8072		2;2 ^a	20	41	42;50 ^a	258	330	40;40;37
Effluent from Chlorinator	8052	ND;ND ^a	20	60	17;19 ^a	258	72	47;51
Effluent from Carbon Treatment	8062	1	-	-	21	-	-	-
Duplicate 8062		1;ND ^a	20	15	10;14 ^a	258	47	40;41;42
C. Reverse Osmosis Treatment								
Influent to Treatment	8073	Sample Bottle Broken in Transit			-	-	-	-
Brine Sample	8083	2;2 ^a	NA ^c	NA ^c	14;12 ^a	NA	NA	69
Effluent from Treatment	8093	3	-	-	24	-	-	-
Duplicate 8093		4;1	20	20	96;82 ^a	258	81	47;46;47
Carriage Water	8003	ND	NA ^c	NA ^c	10	NA	NA	NA ^c

a. Duplicate injections.

b. Not detected.

c. Not required by protocol.

d. Surrogate spikes not used.

verification. Priority pollutant spike recoveries varied considerably. The variation in recoveries is attributed in part to contamination of the samples by laboratory glassware. For example, the very high recovery for sample 8072 is obviously due to contamination. More consistent recovery values were obtained for the surrogate spiked samples. A very interesting surrogate recovery value was obtained for the reverse osmosis brine sample. Whereas the recoveries ranged from 40 to 50% for the other samples, a >60% recovery was obtained for the brine sample. This result is attributed to the high ionic strength of the sample which increase extraction efficiency.

4.3 TOTAL CYANIDE RESULTS

Total cyanide results shown in Table 4-4 were determined by the EPA method for total cyanides contained in the manual of Methods for Chemical Analysis of Water and Wastes, EPA-625/6-74-003, 1974. The samples were field stabilized by the addition of sodium hydroxide and in the case of the chlorinator samples, by the addition of ascorbic acid. The results presented in Table 4-1 shown that the cyanide concentrations in the raw influents were about the same for the first and third sampling periods whereas the cyanide concentration dropped significantly for the second treatment sampling.

All of the process treatments evaluated decreased the cyanide concentration of the waste stream. Based upon cyanide concentration in the final effluent, ozonation is the most effective process for cyanide reduction and the alkaline chlorination/activated carbon treatment is the second most effective process. Reverse osmosis appears to be the least effective treatment. However, based on percent reduction of pollutant loading (i.e., $\frac{\text{raw influent} - \text{effluent}}{\text{raw influent}} \times 100$) across the treatment unit, the alkaline chlorination and reverse osmosis processes are about equally effective. As would be expected, the reverse osmosis brine sample contained the highest cyanide level.

Table 4-4 Total Cyanide Results

Treatment/Sample Description	TRW Sample Number	Total Cyanide, $\mu\text{g}/\text{l}$
A. O_3 Treatment		
Influent to Treatment	8011B	208;278
Effluent from Treatment	8012B	<0.2
Carriage Water	8013B	5
B. Alkaline Chlorination/ Activated Charcoal Treatment		
Influent to Treatment	8072B	90
Effluent from Chlorinator	8052B	24
Effluent from Carbon Treatment	8062B	6
C. Reverse Osmosis Treatment		
Influent to Treatment	8073B	298
Brine Sample	8083B	1210
Effluent from Treatment	8093B	15
Carriage Water	8003B	1.8

APPENDIX A-3.2
RESEARCH TRIANGLE INSTITUTE
PRIORITY POLLUTANT RUN NO. 15 ANALYSIS

Ecological, Mutagenicity, and Organic Test Results, Run 15

Samples taken during Run 15 (raw feed, chlorination effluent, and activated carbon effluent) were analyzed by Research Triangle Institute for ecological effects, mutagenicity, toxicity, and general organic content. Results obtained indicate that the wastewaters tested have no significant ecological or mutagenic potentials.

The ecological tests measure the effect of the water components on *Daphnia Pulex*, a representative aquatic organism. The test is a static-bioassay (using the sample at various dilutions) and is scored by determining the number of dead or affected organism after a specified period, i.e., 24 or 48 hours.

Mutagenicity was determined using the Ames Salmonella tests procedure with five bacteria strains (TA 98, TA 100, TA 1537, TA 1538, TA 1535). The test is scored by comparing the number of revertant bacteria colonies to the number of spontaneous revertants.

Toxicity was determined using the above salmonella strains. The test is conducted by plating bacteria colonies on a petri dish, exposing these colonies to the sample in question, washing the test medium and introducing a growth medium. The test is scored by comparing the number of colonies that form to the number originally plated.

Organic content is determined by extracting the sample at pH2 and pH12 with methylene chloride. After determining the quantity of organic extracted, the sample is fractionated (assuming sufficient organic was extracted) by liquid chromatography and each fraction analyzed by infrared and low resolution mass spectroscopy.

Summary of Results

Ecological Tests

The results of the ecological tests (TABLE 1) show that each treatment step resulted in an improvement in the water quality. The 38 percent value found for the raw wastewater indicates it is only moderately toxic. The complete test data is shown in TABLE 2.

TABLE 1. Ecological Results Summary, RUN 15

Water Type	48 hour LC ₅₀ , Percent ⁶
Raw water	38
After chlorination ^a	65
After Activated Carbon ^a	100

a Sodium thiosulfate had been added for chlorine neutralization

b Percent wastewater in diluted sample that kills 50 percent of the organisms after 48 hour exposure.

TABLE 2. STATIC BIOASSAY DATA, RUN 15

Source Blast Furnace Date Obtained March 15, 1978
 Investigator Edward Monnig / Rosemary Batten
 Date/Time Initiated 4/26/79 1:00 PM Date/Time Completed 4/28/79 2:00 PM
 Source of Dilution Water Carbon Filtered Durham City H₂O
 Test Species Daphnia pulex Temperature Range 21.0 - 22.0°C
 Number Individuals Per Percent Waste 10
 Calculations and Report Reviewed by C. W. Westbrook
 Sample ID Number Raw Water (39313) Run # 15

Start

Percent waste	10%	18%	32%	56%	100%	Control
DO, mg/l	8.2		8.5		8.4	8.7
Temperature, °C	21.5		21.5		1.5	22
pH	7.35		7.7		7.5	6.6
Specific conductance, µmhos	300		750		1900	105

24 hours

Number surviving	9	10	7	3	0	9
% survival	90	100	70	30	0	90
DO, mg/l	8.2		8.5		8.4	8.3
Temperature, °C	22		22		22	21.5
pH	7.4		7.7		7.6	6.7

48 hours

Number surviving	9	10	6	3	0	9
% survival	90	100	60	30	0	90
DO, mg/l	8.2		8.0		8.2	8.2
Temperature, °C	21		21		21	21

48 HOUR LC₅₀ = 38%

Calculations and Report by C. W. Westbrook

Sample ID Number Chlorination Effluent (39315)

Start						
Percent waste	10%	18%	37%	56%	100%	Control
DO, mg/l	8.5		8.8		9.0	8.7
Temperature, °C	21.5		21.5		22	22
PH	7.6		7.6		7.7	6.6
Specific conductance, μ mhos	460		1300		3450	105
24 hours						
Number surviving	9	10	10	10	0	9
% survival	90	100	100	100	0	90
DO, mg/l	8.5		8.3		8.2	8.3
Temperature, °C	21.5		21.5		22	21.5
PH	7.5		7.6		7.6	6.7
48 hours						
Number surviving	9	10	10	7	0	9
% survival	90	100	100	70	0	90
DO, mg/l	8.2		8.3		8.2	8.2
Temperature, °C	20.5		20.5		20.5	21

48 HOUR $1C_{50}$ = 65%

Calculations and Report by C. W. Westbrook

Sample ID Number Activated carbon column effluent (39317)

Start						
Percent waste	10%	18%	37%	56%	100%	Control
DO, mg/l	8.3		8.5		8.4	8.7
Temperature, °C	21.5		21		21	22
PH	6.9		7.05		7.3	6.6
Specific conductance, μ mhos	390		1150		3100	105
24 hours						
Number surviving	10	10	7	8	6	9
% survival	100	100	70	80	60	90
DO, mg/l	8.2		8.5		8.7	8.3
Temperature, °C	21.5		21.5		21.5	21.5
pH	7.0		7.1		7.2	6.7
48 hours						
Number surviving	10	10	7	7	5	9
% survival	100	100	70	70	50	90
DO, mg/l	8.0		8.4		8.2	8.2
Temperature, °C	20.5		20.5		20.5	21

48 hour $LC_{50} = 100\%$

Mutagenicity/Toxicity Tests

The samples were tested for mutagenicity and toxicity in the as received condition and after concentrating by a factor of ten. The bacteria strains were treated with different concentrations of each sample to determine any dose-response effects. All tests in this series were negative indicating no mutagenic or toxic effects. The data obtained is shown in Tables 3 and 4.

Organic Content

Each of the three samples was extracted at pH2 and pH12 with methylene chloride. None of the samples contained enough organic matter to warrant further organic category characterization. The results (Table 5) do show that alkaline chlorination removed essentially all of the organic material.

TABLE 3. MUTAGENIC RESPONSE DATA, RUN 15

Sample type	Dose Applied	TA98		TA100		TA1537		TA1538		TA1535	
		-MA	+MA	-MA	+MA	-MA	+MA	-MA	+MA	-MA	+MA
Raw Influent 39314	conc	1.23	0.76	1.33	1.32	1.05	1.39	0.70	1.01	1.08	0.91
	1:2	1.16	0.93	1.35	1.21	0.87	1.35	0.56	1.01	1.33	1.10
	1:5	1.18	0.95	1.38	1.29	1.21	1.21	0.67	0.84	1.62	0.93
	1:10	0.96	0.92	1.45	1.29	0.91	1.60	0.69	0.83	1.58	0.90
Chlorine Con- tact Effluent 39316	conc	1.12	1.04	1.19	1.23	1.09	1.56	0.92	0.76	0.86	1.06
	1:2	0.99	0.97	1.41	1.36	0.73	1.33	0.83	1.34	1.03	1.07
	1:5	0.99	1.04	1.22	1.27	1.09	1.25	0.73	1.12	0.96	0.88
	1:10	1.08	0.95	1.27	1.35	1.01	1.11	1.01	1.39	1.09	1.13
Carbon Filter Effluent 39318	conc	1.28	0.99	1.53	1.37						
	1:2	2.01	1.67	1.39	1.40						
	1:5	1.32	1.03	--	--	0.01	0.89	0.76	1.04	1.12	1.15
	1:10	1.29	1.00	1.44	1.39	0.81	0.77	0.73	0.85	1.21	1.09
Raw Influent 39314 concentrated	10:1	1.54	0.80	1.12	1.01	0.99	1.31	0.84	1.03	0.92	0.90
	5:1	1.17	0.87	0.94	0.99	0.97	0.88	0.62	0.91	0.89	1.03
	2.5:1	1.36	1.00	1.03	0.97	0.82	1.12	0.66	0.86	0.97	0.96
	1:10	1.24	1.02	1.14	0.94	0.93	1.20	0.43	1.09	1.01	0.94
	1:100	0.90	0.88	1.24	0.90	0.79	1.50	0.62	1.01	0.90	0.86
Chlorine Con- tent Effluent 39316 concentrated	10:1	1.18	0.76	1.23	1.30	0.90	1.70	0.76	0.62	1.10	0.96
	5:1	1.05	0.86	1.22	1.25	0.95	1.31	0.51	0.84	0.90	0.92
	25:1	1.06	0.98	1.17	1.30	0.88	0.90	0.58	0.64	1.05	0.91
	1:10	1.17	1.07	1.07	1.07	0.90	1.22	0.72	0.91	0.72	0.86
	1:100	1.07	1.02	1.08	1.16	0.71	1.08	0.60	0.78	0.99	1.03
Carbon Filter Effluent 39318 concentrated	10:1	1.21	0.91	1.10	1.06	0.69	1.05	0.78	0.66	0.99	1.14
	5:1	1.31	1.11	1.25	1.04	0.93	1.35	0.95	1.09	1.07	1.12
	2.5:1	1.19	0.99	0.94	1.19	0.97	1.10	0.47	0.74	0.88	1.06
	1:10	1.23	1.09	1.38	1.08	1.08	0.92	0.63	0.64	0.90	0.83
	1:100	1.41	1.06	1.18	1.33	0.92	0.93	0.35	0.62	0.97	1.06

Positive Control	47.06	32.72	16.89	23.67	toxic	11.12	71.11	54.13	25.64	8.32
Solvent Control	1.00	1.00	1.00	1.00	1.00	1.00		1.00	1.00	1.00

- a. bacteria strain
- b. -MA is uninduced activity, +MA is induced activity
- c. Revertants per plate divided by spontaneous revertants.

TABLE 4. Toxicity Response Data, Run 15

Sample type	Dose Applied	TA98 -MA+MA	TA100 -MA+MA	TA1537 -MA-MA	TA1538 -MA+MA	TA1538 -MA+MA
Raw Influent 39314	conc 1:2 1:5 1:10	1.71 ^c 1.90 1.29 1.12	1.05 0.97 1.05 1.11	1.22 1.43 1.11 1.47	1.22 1.63 1.12 1.64	1.04 0.80 0.96 .84
Chlorine Con- tact Effluent 39316	conc 1:2 1:5 1:10	1.34 1.67 1.19 1.22	1.05 1.02 1.07 1.02	1.36 1.32 1.32 1.43	1.01 1.84 1.07 1.62	1.11 0.96 1.06 0.97
Carbon Filter Effluent 39318	conc 1:2 1:5 1:10	1.63 2.09 2.30 1.83	1.03 1.13 0.98 1.02	1.21 1.36 1.34 1.15	1.27 1.68 -- 1.38	0.96 1.18 1.13 1.33
Raw Influent 39314	10:1 5:1 2.5:1 1:10 1:100	0.73 0.79 0.45 1.15 0.52	1.05 1.05 0.92 1.01 0.99	1.13 0.95 1.03 1.11 1.00	1.21 1.48 1.28 1.28 1.34	0.85 1.14 1.03 1.12 1.08
Chlorine Con- tent Effluent 39316	10:1 5:1 2.5:1 1:10 1:100	1.10 0.63 1.14 2.25 0.85	1.13 1.01 0.95 1.15 1.03	0.91 1.19 0.98 0.89 0.85	1.17 1.23 1.19 1.33 1.25	0.80 1.16 1.16 1.08 1.07
Carbon Filter Effluent 39318	10:1 5:1 2.5:1 1:10 1:100	2.05 2.31 2.08 2.07 0.79	1.00 1.08 0.99 1.01 1.03	1.19 1.07 1.20 1.31 1.06	1.37 1.27 1.31 1.30 1.24	1.11 1.16 1.26 1.03 0.98
						0.58 0.75

Solvent Control 1.00 1.00 1.00 1.00 1.00 1.00 1.00

- a. bacteria strain
- b. -MA is uninduced activity, +MA is induced activity
- c. colony forming units per plate divided by colony forming units plated.

TABLE 5. ORGANIC IN RUN 15 SAMPLES

<u>SAMPLE</u>	<u>ORGANIC CONTENT, mg/l</u>
Raw Wastewater (39313)	2.5
Alkaline Chlorination Effluent	0.0
Activated Carbon Effluent	0.0

TABLE A-4. CHILTON FACTORS USED FOR COST ESTIMATE
OF THE VARIOUS TREATMENT PROCESSES

Item number	Item description	Alkaline chlorination	Chemical clarification	Dual media filtration	Reverse osmosis	Ozonation	Activated carbon
(1)	Delivered equipment cost						
(2)	Equipment installation	.4(1)	.4(1)	.4(1)	.4(1)	.4(1)	.4(1)
(3)	Installed equipment cost	(1)+(2)	(1)+(2)	(1)+(2)	(1)+(2)	(1)+(2)	(1)+(2)
(4)	Process piping	.3(3)	.3(3)	.3(3)	.3(3)	.3(3)	.3(3)
(5)	Instrumentation	.05(3)	.05(3)	.05(3)	.05(3)	.05(3)	.05(3)
(6)	Capital equipment total	(3)+(4)+(5)	(3)+(4)+(5)	(3)+(4)+(5)	(3)+(4)+(5)	(3)+(4)+(5)	(3)+(4)+(5)
(7)	Building & Site Development - outdoor - outdoor/indoor - indoor	.8(6)	.3(6)	.8(6)	.8(6)	.6(6)	.8(6)
(8)	Auxiliaries (Electrical Item Power)	.2(6)	.2(6)	.2(6)	.2(6)	.2(6)	.2(6)
(9)	Other	.1(6)	.1(6)	.1(6)	.1(6)	.1(6)	.1(6)
(10)	Facilities total	(7)+(8)+(9)	(7)+(8)+(9)	(7)+(8)+(9)	(7)+(8)+(9)	(7)+(8)+(9)	(7)+(8)+(9)
(11)	Total physical plant costs	(6)+(10)	(6)+(10)	(6)+(10)	(6)+(10)	(6)+(10)	(6)+(10)
(12)	Engineering and construction	.25(11)	.25(11)	.25(11)	.25(11)	.25(11)	.25(11)
(13)	Contingency & contractor's fee	.25(11)	.25(11)	.25(11)	.25(11)	.25(11)	.25(11)
(14)	Facilities & fees	(10)+(12)+(13)	(10)+(12)+(13)	(10)+(12)+(13)	(10)+(12)+(13)	(10)+(12)+(13)	(10)+(12)+(13)
(15)	Total Plant Cost	(6)+(14)	(6)+(14)	(6)+(14)	(6)+(14)	(6)+(14)	(6)+(14)

(n) Indicates item number being multiplied or added.

TABLE A-5. TYPICAL COST ESTIMATE UTILIZING CHILTON FACTORS

Item Number	Item Description	Cost ^a	Comments
(1)	Delivered equipment cost		--
(2)	Equipment installation .4x(1)	\$100,000 40,000	Factor used is at low end of range because some equipment is skid mounted.
(3)	Installed equipment cost		--
(4)	Process piping .3x(3)	<u>\$140,000</u> 42,000	Factor used is at low end of range for fluid piping because some equipment is prepip.
(5)	Instrumentation .05x(3)	7,000	Factor represents minor instrumentation.
(6)	Capital equipment total (3+4+5)	<u>\$189,000</u>	--
(7)	Buildings & site development .8x(6)	\$151,000	Factor is at midrange for indoor installation.
(8)	Auxiliaries (Electrical Power) .2x(6)	38,000	Factor is at low midrange for a major addition.
(9)	Other .1x(6)	19,000	Factor is near low end of range.
(10)	Facilities Total (7+8+9)	<u>\$208,000</u>	--
(11)	Total Physical Plant Costs (6+10)	\$397,000	Capital equipment plus facilities.
(12)	Engineering and construction fee .25(11)	99,000	Factor is for simple to moderate complexity.
(13)	Contingency & contractor's fee .25(11)	99,000	Factor is midrange for processes subject to change.
(14)	Facilities & fees (10+12+13)	\$406,000	--
(15)	Total Plant Cost (6+14)	\$595,000	--

(n) Indicates Item Number Being Multiplied.

a Rounded to the nearest thousand.

TABLE B-1. ALKALINE CHLORINATION COMPOSITE SAMPLE DATA

	pH	SS	VSS	TOC	800	OCG	CNA	Phenol NH ₃	TKN	S ⁻	F ⁻	TDS	Cl ⁻	NO ₂ /NO ₃	Color	TP
Run #1																
Influent	7.8															
Effluent	7.4							46								
								1.3								
Run #2																
Influent	7.8	55	7	26	--	--	<0.12	0.029	38.3	40.3	<0.2	11.5	1238	410	0.06	15
Effluent	7.3	8	1	22	--	--	<0.01	0.014	1.7	2.0	<0.2	10.5	2820	1375	2.30	10
																<0.02
Run #3																
Influent	7.8															
Effluent	7.4							37.5								
								1.0								
Run #4																
Influent	7.7	75	17	28	--	--	<0.01	0.020	35.9	36.5	<0.2	14.8	1125	385	0.10	15
Effluent	6.7	11	8	15	--	--	0.11	0.012	7.4	7.9	<0.2	14.2	2527	1270	0.74	15
																<0.05
Run #5																
Influent	7.8	870	360	180	--	16	0.17	0.011	36.3	39.8	<0.2	21.4	1474	465	0.1	15
Effluent	7.2	33	13	19	--	<1	<0.01	0.004	4.8	3.0	<0.2	19.2	3053	1520	1.42	15
																0.02
Run #6																
Influent	7.8	73	8	35	--	--	0.19	0.024	32.9	35.3	<0.2	13.5	1120	390	0.06	15
Effluent	7.3	16	4	12	--	--	<0.01	0.014	0.48	1.6	<0.2	12.5	2280	1100	1.96	20
																--
Run #7																
Influent	7.8	51	11	17	21	--	0.10	0.020	29.0	31.4	0.25	13.1	1100	380	0.09	5/10
Effluent	6.9	16	7	11	4.0	--	<0.01	0.011	0.67	2.6	<0.20	10.9	1810	835	0.95	5/10
																<0.02
Run #8																
Influent	7.5	85	15	33	--	--	0.02	0.027	29.2	31.5	0.29	11.6	1180	400	0.11	20/30
Effluent	7.2	16	6	10	--	--	<0.01	0.004	0.67	1.2	<0.20	11.0	2350	1120	2.20	5/10
																<0.02
Run #9																
Influent	7.3	66	18	35	--	<1	0.01	0.015	30.4	31.4	<0.2	13.4	1058	335	--	5/10
Effluent	6.3	--	--	13	--	1	<0.01	0.008	10.3	11.8	<0.2	12.4	2162	120	0.1	10
																0.03
Run #10																
Influent	7.7	62	23	35	--	--	0.02	0.017	32.4	33.5	<0.2	13.7	1001	475	0.1	15
Effluent	6.7	25	10	15	--	--	<0.01	0.009	5.2	5.6	<0.2	13.4	2142	1160	--	15
																<0.04

NOTE: All parameters in mg/l except pH (units) and color (APHA units)

(continued)

TABLE B-1. (continued).

	pH	SS	VSS	TOC	BOD	O&G	CNA	Phenol	NH ₃	TKN	S ²⁻	F ⁻	TDS	Cl ⁻	NO ₂ /NO ₃	Color	TP	Fe
Run #11																		
Influent	7.9	3480	630	1340	<15	--	0.18	0.044	17.0	46.6	<0.2	13.7	1433	470	0.08	10	2.52	420
Effluent	7.2	35	18	26	17.1	--	<0.01	0.013	6.95	8.5	<0.2	13.6	2689	1390	0.59	10	<0.02	3.1
Run #12																		
Influent	7.9	3595	335	620	<15	--	0.43	0.030	32.2	48.2	<0.2	16.2	1395	460	0.11	20	1.91	
Effluent	6.8	33	10	20	29	--	<0.01	0.009	10.9	13.4	<0.2	15.9	2456	1300	0.57	15	0.06	
Run #13																		
Influent	7.9	1810	315	280	<15	11	0.10	0.077	28.6	39.9	<0.2	14.6	1206	420	0.13	15	1.19	180
Effluent	7.0	29	10	14	14.5	4	<0.01	0.009	5.7	7.8	<0.2	13.9	2028	1100	0.46	10	<0.03	1.0
Run #14																		
Influent	7.7	49	9	13	--	15	0.12	0.029	32.8	33.3	<0.1	12.5	1080	390	0.08	0/5		
Effluent	7.2	7	6	15	--	3	0.02	0.386	5.2	5.9	<0.1	14.4	1840	950	0.59	0/5		
Run #15																		
Influent	7.3	26	6	35	--	17	0.34	0.019	33.3	37.9	<0.2	21.4	1210	438	0.07	0/5	0.07	4.8
Effluent	6.4	29	11	32	--	3	<0.01	<0.001	7.35	9.9	<0.2	20.8	2020	1025	<0.01	5/10	0.51	3.3
Run #16																		
Influent	7.8	95	18	45	--	--	0.23	0.026	34.0	40.2	<0.2	18.9	1160	425	0.06	5	0.08	--
Effluent	7.4	25	7	46	--	--	<0.01	0.005	6.15	9.5	<0.2	18.0	2025	978	<0.01	0	0.40	--
Run #17																		
Influent	7.9	101	22	62	10.4	1	0.28	0.017	38.4	38.9	<0.2	16.9	1255	400	0.08	0	0.06	--
Effluent	7.4	37	11	32	13.8	2	<0.01	0.005	11.9	14.8	<0.2	16.8	2155	1010	0.46	0	<0.01	--
Run #18																		
Influent	8.0	122	22	45	18.2	--	0.23	0.020	37.8	42.3	<0.2	18.0	1280	395	0.07	0/5	0.10	--
Effluent	7.1	36	9	25	20	--	<0.01	0.005	4.8	6.7	<0.2	16.9	2200	1160	0.66	0	0.02	--
Run #19																		
Influent	7.5								34.5									
Effluent	6.9								3.1									
Run #20																		
Influent	7.5								32.0									
Effluent	6.7								16.0									

NOTE: All parameters in mg/l except pH (units) and color (APHA units)

TABLE B-2. ALKALINE CHLORINATION DATA SUMMARY - RAW WASTEWATER

Run no.	Contact time, min.	pH	Suspended solids, mg/l	Cyanide, mg/l	Ammonia, mg/l	NaOH used gal/1000 gal treated	HCl used gal/1000 gal treated	Cl ₂ Residual mg/l
Influent	1							
CN tank	(3 gpm)	7.7	--	--	46			0
NH ₃ tank	20	--	--	--	--	2.48 @ 10%	4.0 @ 1.0%	134.7
Clarifier	24	7.5	--	--	0.8			200.6
Effluent	240	7.4	--	--	1.3			0
Influent	2							
CN tank	(3 gpm)	7.8	55	0.12	38.3			0
NH ₃ tank	20	10.5	--	--	--	2.7 @ 10%	4.7 @ 1%	--
Clarifier	24	7.3	--	--	--			53
Effluent	240	7.3	8	<0.01	1.7			124
Influent	3							
CN tank	(3 gpm)	7.8	--	--	37.5			0
NH ₃ tank	20	--	--	--	--	2.72 @ 10%	5.21 @ 1%	0
Clarifier	24	7.4	--	--	1.0			177
Effluent	240	7.4	--	--	1.0			113
Influent	4							
CN tank	(3 gpm)	7.7	75	<0.01	35.9			0
NH ₃ tank	20	10.2	366	<0.01	19.3	2.5 @ 18%	9.9 @ 1%	--
Clarifier	24	6.9	192	<0.01	14.8			--
Effluent	240	6.7	11	0.11	7.4			57.6
Influent	5							
CN tank	(3 gpm)	7.8	870	0.17	36.3			0
NH ₃ tank	20	--	855	<0.01	11.2	3.9 @ 18%	8.2 @ 1%	141
Clarifier	24	--	1320	<0.01	2.5			17.7
Effluent	240	7.2	33	<0.01	4.8			23
Influent	6							
CN tank	(5 gpm)	7.8	73	0.19	32.9			0
NH ₃ tank	12	10.75	--	--	--	3.95 @ 10%	7.31 @ 1%	--
Clarifier	15	7.1	137	0.10	25.0			7.1
Effluent	144	7.3	16	<0.01	0.48			42

Note: gpm x 3.785 = l/min
gal. x 3.785 = liters

continued

TABLE B-2. (continued)

Run no.	Contact time, min.	pH	Suspended solids, mg/l	Cyanide, mg/l	Ammonia, mg/l	NaOH used gal/1000 gal treated	HCl used gal/1000 gal treated	Cl ₂ Residual mg/l
7								
Influent		7.8	51	0.10	29.0			0
CN tank	12	10.75	--	--	--	2.37 @ 10%	1.40 @ 1%	--
NH ₃ tank	15	6.8	220	<0.01	0.42			39
Clarifier Effluent	144	6.9	16	<0.01	0.67			13.2
8								
Influent		--	85	0.02	29.2			0
CN tank	12	10.75	772	<0.01	4.4	1.83 @ 10%	3.5 @ 1%	--
NH ₃ tank	15	7.2	264	<0.01	0.62			145
Clarifier Effluent	144	7.2	16	<0.01	0.67			133
9								
Influent		--	66	0.01	30.4			0
CN tank	12	11.0	545	<0.01	18.7	2.0 @ 18%	6.1 @ 1%	--
NH ₃ tank	15	6.7	152	<0.01	11.6			--
Clarifier Effluent	144	6.3	--	<0.01	10.3			103
10								
Influent		--	62	0.02	32.4			0
CN tank	12	10.8	482	<0.01	--	2.1 @ 18%	5.2 @ 1%	--
NH ₃ tank	15	6.7	200	<0.01	--			--
Clarifier Effluent	144	6.7	25	<0.01	5.2			60.3
11								
Influent		--	3480	0.18	46.0			0
CN tank	12	11.0	740	0.02	32.4	3.1 @ 18%	7.4 @ 1%	132.9
NH ₃ tank	15	7.3	1455	0.12	25.6			139.6
Clarifier Effluent	144	7.2	35	<0.01	6.95			47.5
12								
Influent		--	3595	0.43	32.2			0
CN tank	12	11.1	3020	<0.01	3.4	2.1 @ 18%	5.3 @ 1%	93
NH ₃ tank	15	7.0	2070	0.30	5.1			124
Clarifier Effluent	144	6.8	33	<0.01	10.9			81

Note: gpm x 3.785 = l/min
gal. x 3.785 = liters

continued

TABLE B-2. (continued)

Run no.	Contact time, min.	pH	Suspended solids, mg/l	Cyanide, mg/l	Ammonia, mg/l	NaOH used gal/1000 gal treated	HCl used gal/1000 gal treated	Cl ₂ Residual mg/l
Influent	13	--	1810	0.10	28.6			0
CN tank	(5 gpm)	12	3560	<0.01	6.8	2.5 @ 18%	4.2 @ 1%	145
NH ₃ tank	15	6.9	3000	<0.01	6.8			73
Clarifier Effluent	144	7.0	29	<0.01	5.7			79.3
Influent	14	--	49	0.12	32.8			0
CN tank	(6.4 gpm)	7.7	--	--	--	1.81 @ 10%	2.51 @ 1%	--
NH ₃ tank	9.4	--	--	--	--			--
Clarifier Effluent	11.3	--	--	--	--			--
Influent	15	7.2	7	0.02	5.2			--
CN tank	--	7.3	26	0.34	33.3	1.5 @ 18%	4.0 @ 1%	0
NH ₃ tank	7.1	11.1	750	<0.01	2.3			168
Clarifier Effluent	8.5	7.9	380	0.09	3.2			117
Influent	16	--	29	<0.01	7.35			113
CN tank	(8.5 gpm)	7.8	95	0.23	34.0			0
NH ₃ tank	7.1	9.7	750	<0.01	5.5	1.2 @ 18%	4.3 @ 1%	56.8
Clarifier Effluent	8.6	7.3	290	<0.01	6.8			113
Influent	17	7.4	25	<0.01	6.15			106
CN tank	(8.5 gpm)	7.9	101	0.28	38.4			0
NH ₃ tank	7.1	11.8	785	<0.01	14.2	2.1 @ 18%	5.0 @ 1%	184
Clarifier Effluent	8.6	7.7	450	<0.01	11.4			149
Influent	18	7.4	37	<0.01	11.9			99.2
CN tank	--	8.0	122	0.23	37.8			0
NH ₃ tank	(8.5 gpm)	7.1	975	<0.01	6.2	2.0 @ 18%	4.3 @ 1%	246
Clarifier Effluent	8.6	7.7	490	<0.01	8.3			142
Influent	18	7.10	36	<0.01	4.8			65.8

Note: gpm x 3.785 = l/min
gal. x 3.785 = liters

continued

TABLE B-2. (continued)

Run no.	time, min.	pH	Suspended solids, mg/L	Cyanide, mg/L	Ammonia, mg/L	NaOH used gal/1000 gal treated	HCl used gal/1000 gal treated	Cl ₂ Residual mg/L
19	--	7.4	--	--	34.5	--	--	0
CN tank	7.1	10.3	--	--	--	1.9 @ 18%	4.8 @ 1%	--
NH ₃ tank	8.6	7.0	--	--	--	--	--	10.16
Clarifier	85.7	6.9	--	--	3.1	--	--	141
Effluent								
20	--	7.5	--	--	32	--	--	0
CN tank	6.8	10.4	--	--	12	1.8 @ 18%	4.6 @ 1%	83
NH ₃ tank	8.2	6.9	--	--	21	--	--	80
Clarifier	82.8	6.7	--	--	16	--	--	86
Effluent								

Note: gpm x 3.785 = L/min
gal. x 3.785 = liters

TABLE B-3. ALKALINE CHLORINATION COMPOSITE SAMPLES DATA
REVERSE OSMOSIS BRINE

	pH	SS	VSS	TOC	CNA	Phenol	NH ₃	TKN	S ⁻	F ⁻	TDS	Cl ⁻
Run #1												
Influent	6.1	--	--	10	--	0.659	87	--	--	--	5616	--
Effluent	6.8	88	14	7	<0.01	0.002	1.8	--	<0.2	21.3	--	--
Run #2												
Influent	6.2	--	--	8	<0.01	0.023	90	81.0	<0.2	18.9	5690	2250
Effluent	6.8	54	10	5	<0.01	0.010	4.2	--	<0.2	20.1	7996	3650
Run #3												
Influent	6.15	--	--	8	<0.01	0.023	90	81.0	<0.2	18.9	5690	2250
Effluent	6.35	--	--	5	--	--	2.4	3.9	--	--	7424	3300
Run #4												
Influent	6.1	--	--	10	<0.01	0.022	123	120	<0.2	35.2	7655	3050
Effluent	7.0	1	3	5	0.02	0.010	25	24.0	<0.2	23.5	7140	3450
Run #5												
Influent	6.7	--	--	10	<0.01	0.022	123	120	<0.2	35.2	7655	3050
Effluent	5.8	42	5	7	<0.01	0.015	<0.05	1.1	<0.2	21.9	9482	4550
Run #6												
Influent	6.3	--	--	16	0.08	0.011	89	77.0	<0.2	9.37	5500	2325
Effluent	7.85	--	--	12	<0.01	0.005	3.25	3.0	--	7.03	8106	3750
Run #7												
Influent	6.35	--	--	16	0.08	0.011	89	77.0	<0.2	9.37	5500	2325
Effluent	6.95	--	(No effluent sample)									
Run #8												
Influent	6.45	--	--	10	<0.01	0.016	73	66.6	<0.2	11.44	4310	1300
Effluent	6.5	--	--	13	0.03	0.003	23.8	--	--	10.05	7061	3300
Run #9												
Influent	7.35	--	--	10	<0.01	0.016	73	66.6	<0.2	11.44	4310	1350
Effluent	6.6	34	5	15	<0.01	0.005	16.8	16.0	--	10.56	6105	2780

TABLE B-4. ALKALINE CHLORINATION DATA SUMMARY - REVERSE OSMOSIS BRINE

Run no.	Contact time, min.	pH	Suspended solids, mg/l	Cyanide, mg/l	Ammonia, mg/l	NaOH used gal/1000 gal treated	HCl used gal/1000 gal treated	Cl ₂ Residual mg/l
Influent	1	--	6.1	--	8.7	--	--	0
CN tank	(1.0 gpm)	30.0	11.5	--	16.5	--	--	172
NH ₃ tank	10.0	6.55	--	<0.01	8.5	--	--	132
Clarifier Effluent	360	6.8	88	<0.01	1.8	--	--	18
Influent	2	--	6.2	--	90	--	--	0
CN tank	(2.5 gpm)	12.0	10.5	<0.01	<0.05	4.9 @ 18%	12.2 @ 1%	29.5
NH ₃ tank	4.0	6.5	143	--	3.25	--	--	26.5
Clarifier Effluent	144	6.8	54	<0.01	4.2	--	--	14.1
Influent	3	--	6.15	--	90	--	--	0
CN tank	(2.5 gpm)	12.0	9.8	<0.01	17.3	3.6 @ 18%	9.8 @ 1%	9.7
NH ₃ tank	4.0	6.75	410	--	5.25	--	--	14.1
Clarifier Effluent	144	6.35	--	--	2.4	--	--	6.2
Influent	4	--	6.1	--	<0.01	123	--	0
CN tank	(2.5 gpm)	12.0	11.0	--	--	--	--	31.9
NH ₃ tank	4.0	6.75	--	--	--	--	--	29.2
Clarifier Effluent	144	7.0	1	0.02	25	--	--	55.2
Influent	5	--	6.7	--	<0.01	123	--	0
CN tank	(5.0 gpm)	6.0	11.3	--	<0.01	--	--	--
NH ₃ tank	2.0	6.8	--	--	<0.05	--	--	--
Clarifier Effluent	72	5.8	42	<0.01	<0.05	--	--	283.6
Influent	6	--	6.3	--	89	--	--	0
CN tank	(2.5 gpm)	12.0	10.3	0.08	13.8	7.6 @ 18%	11.4 @ 1%	77.9
NH ₃ tank	4.0	8.8	834	--	8.5	--	--	127.6
Clarifier Effluent	144	7.85	--	<0.01	3.25	--	--	60.2

Note: gpm x 3.785 = l/min
gal. x 3.785 = liters

continued

TABLE B-4. (continued)

Run no.	Contact time, min.	pH	Suspended solids, mg/l	Cyanide, mg/l	Ammonia, mg/l	NaOH used gal/1000 gal treated	HCl used gal/1000 gal treated	Cl ₂ Residual mg/l
7	--	6.35	--	0.08	89			0
CN tank	6.0	10.8	1758	0.11	20	6.67 @ 18%	16	20.0
NH ₃ tank	2.0	6.75	330	--	20.5	@ 18%	@ 1%	23.0
Clarifier Effluent	72	6.95	--	--	--			8.0
8	--	6.45	--	<0.01	73			0
CN tank	6.0	10.2	2051	0.02	22.3	4.7 @ 18%	11.7 @ 1%	172
NH ₃ tank	2.0	6.85	881	--	23.8	@ 18%	@ 1%	213
Clarifier Effluent	72	6.5	--	0.03	23.8			115
9	--	7.35	--	<0.01	73			0
CN tank	5.0	9.9	2388	0.02	0.75	2.6 @ 18%	10.3 @ 1%	28
NH ₃ tank	1.7	6.8	1366	--	3.0	@ 18%	@ 1%	70.9
Clarifier Effluent	60	6.6	34	<0.01	15.8			62

Note: gpm x 3.785 = l/min
gal. x 3.785 = liters

TABLE C-1. CHEMICAL CLARIFICATION DATA FROM BLAST FURNACE WASTEWATER (1.8m diameter clarifier)

Feed Source	Date	Overflow rate gpd/ft ²	Sludge wasting rate ¹ gph	Chemical dose, mg/l FeCl ₃ Alum	Polyelectrolyte dose ² mg/l	SS, mg/l			Turb ³ , FTU			NH ₃ , mg/l			CH ₄ , mg/l			Phenol, mg/l			Fluoride, mg/l			Sulfide, mg/l			pH	
						Inf.	Eff.	grab	Inf.	Eff.	grab	Inf.	Eff.	grab	Inf.	Eff.	grab	Inf.	Eff.	grab	Inf.	Eff.	grab	Inf.	Eff.	grab	Inf.	Eff.
Alkaline Chlorination Effluent	2-11-79	150	10	100	0.5	200	25	36	9	5.2		<0.01			0.009			<0.2			<0.2			<0.2			7.6	7.2
	2-12-79	150	13	100	0.5	192	11	35	5.9	7.4		0.11			0.012			<0.2			<0.2			<0.2			7.2	7.1
	2-13-79	150	11	100	0.5	1320	33	21	12	4.8		<0.01			0.004			<0.2			<0.2			<0.2			7.7	7.3
	1-29-79	250	9	100	0.5				3.0																			
	1-30-79	250	8	100	0.5				3.0																			
	1-31-79	250	8	100	0.5																							
	2-1-79	250	9	100	0.5																							
	2-2-79	250	6	100	0.5																							
	2-3-79	250	19	100	0.5																							
	2-4-79	250	18	100	0.5																							
Raw water	2-5-79	250		100	0.5																							
	2-6-79	250		100	0.5	137	16	25	4.5	0.48		<0.01			0.014			<0.2			<0.2			<0.2			7.8	7.2
	2-7-79	250	11	100	0.5	220	16	25	8.0	0.67		<0.01			0.011			<0.2			<0.2			<0.2			7.8	7.1
	2-8-79	250		100	0.5	264	16	34	14	0.67		<0.01			0.004			<0.2			<0.2			<0.2			7.5	7.2
	2-9-79	250	12	100	0.5																							
	2-10-79	250	10	100	0.5	152				10.3		<0.01			0.008			<0.2			<0.2			<0.2			7.7	6.9
	2-14-79	250	16	100	0.5	1455	35	23	6.4	7.0		<0.01			0.013			<0.2			<0.2			<0.2			7.6	7.0
	2-15-79	400	13	100	0.5	2070	36	31	5.4	10.9		<0.01			0.009			<0.2			<0.2			<0.2			7.7	6.9
	2-16-79	400	20	100	0.5																							
	2-17-79	400	113	100	0.5																							
	2-18-79	400	17	100	0.5																							
	2-19-79	400	14	100	0.5	320	29	40	18	7.4		<0.01			<0.001			<0.2			<0.2			<0.2			7.5	6.7
	2-20-79	400	64	100	0.5	290	25	43	6.8	6.2		<0.01			0.005			<0.2			<0.2			<0.2			7.6	7.0
	2-21-79	400	70	100	0.5	450	37	43	8.5	11.9		<0.01			0.005			<0.2			<0.2			<0.2			7.8	7.0
	2-22-79	400	63	100	0.5	490	36	41	9.1	4.8		<0.01			0.005			<0.2			<0.2			<0.2			7.7	6.9
Raw water	2-27-79	500	7		50																							
	2-28-79	500	19		50																							
	3-1-79	400	7		50	573	42	38	9.0	30.2	40.2				0.023			<0.2			<0.2			<0.2			7.8	7.4
	3-2-79	400	7		50	220	29	42	8.1	31.4	29.6				0.022			<0.2			<0.2			<0.2			7.9	7.3
	3-3-79	400			50	37	21	52	9.1	37.7		0.09			0.022			<0.2			<0.2			<0.2			7.9	7.4

1 Sludge wasting rates are instantaneous values and are not intended to represent average daily rates.

2 Betz 1100.

3 24 hour composite samples.

Conversions: 24.55 x gpd/ft² = m³/m²·d
gph x 3.785 = l/hr

continued

TABLE C-1. (continued)

Feed Source	Date	Overflow rate gpd/ft ²	Sludge wasting rate ¹ gph	Chemical dose, mg/L FeCl ₃	Alum	Polyelectro- lyte dose ² mg/L	SS mg/L		Turb. FTU grab Inf. Eff.	NH ₃ , mg/L		CH ₄ , mg/L		Phenol, mg/L		Fluoride, mg/L		Sulfide, mg/L		pH	
							comp. 3 Inf. Eff.	Inf. Eff.		comp. Inf. Eff.	Inf. Eff.	comp. Inf. Eff.	Inf. Eff.	comp. Inf. Eff.	Inf. Eff.	comp. Inf. Eff.	Inf. Eff.	grab Inf. Eff.	grab Inf. Eff.		
	3-4-79	400	19		50	0.5	161	26	47	21	40.3	0.07	0.015	12.4	<0.2	<0.2	7.9	7.4			
	3-5-79	400	19		50	0.5	101	24	33	7.0	38.2	0.06	0.018	12.7	<0.2	<0.2	7.8	7.2			
	3-6-79	400	19		50	0.5	170	23	38	7.5	36.4	0.03	0.022	13.6	<0.2	<0.2	7.7	7.5			
	3-7-79	400	23		50	0.5	147		46	9.8	40.8	0.06	0.018	8.97	<0.02	<0.02	7.6	7.2			
	3-8-79	400	17		50	0.5	63	27	36	8.4	46.0	0.05	0.019	9.8	<0.02	<0.02	7.7	7.3			
	3-9-79	400	7		50	0.5	84	20	30	7.7	53.4	0.24	0.019	6.44	<0.02	<0.02	7.4	7.3			
	3-10-79	400	17		50	0.5	368	28	33	7.2	53		0.009	9.7	<0.02	<0.02	7.8	7.2			
	3-11-79	400	19		50	0.5	231	21	40	27	38				<0.02	<0.02	7.7	7.5			
	3-12-79	400			50	0.5	98	24			63		0.018	9.3							
	3-13-79	400			50	0.5	99	21			36.5										
	3-14-79	400	19		50	0.5	61		58	40	43		0.033	9.8	<0.02	<0.02	7.8	7.2			
	3-15-79	400	19		50	0.5	315	26	55	24	45	0.32	0.047	8.1	<0.02	<0.02	7.4	7.3			
	3-16-79	400			50	0.5	91	48			15	0.16	0.023	7.5	<0.02	<0.02					
	3-17-79	400	27		100	0.5			27	11					<0.02	<0.02	7.8	7.0			
	3-18-79	400			150	0.5	31	16	44	18	35.5						7.8	7.0			
	3-19-79	400	27		150	0.5	91	26	26	11	30				<0.02	<0.02	7.8	7.0			
	3-20-79	400			150	0.5	229		33	13	31.5	<0.07	0.014	9.2	<0.02	<0.02	7.8	7.0			
	3-21-79	400	26		150	0.5	78	25	35	17	36	<0.01	0.013	9.6	<0.02	<0.02	7.8	8.0			
	3-22-79	400	26		150	0.5	99		35	14	29	0.06	0.018	11.4	<0.02	<0.02	7.8	7.0			
	3-23-79	400			150	0.5	121		35	2.0	52	0.23	0.021	12.7	<0.02	<0.02		6.7			
	3-24-79	400			150	0.5	95	23			31.5	0.68	0.008	8.2	<0.02	<0.02					
	3-25-79	400			150	0.5	84	31	27	4.2	33.5	0.20	0.001	7.16	<0.02	<0.02	7.9	6.7			
	3-26-79	400			150	0.5	97	20	23	3.5	31.5	0.10	0.018	7.03	<0.02	<0.02	7.9	6.8			

¹Sludge wasting rates are instantaneous values and are not intended to represent average daily rates.

²Betz 1100.

324 hour composite samples.

Conversions: 24.55 × gpm/ft² = m³/m²·d
gph × 3.785 = l/hr

TABLE D-1. DUAL MEDIA FILTER - DATA SUMMARY
RAW BLAST FURNACE WASTEWATER FEED

Filtration rate $\frac{L}{min/m^2}$	Run No.	Headloss meters	Run time hrs	Suspended solids, mg/l		Percent solids removal	Solids loading	
				Inf.	Eff.		kg SS/m^2 /m H.L.	lb SS/ft^2 /ft H.L.
122.2	3	7.4	26.7	110	19	83	2.405	0.150
122.2	3	2.5	30.75	128	10	92	10.519	0.656
122.2	18	7.4	56.0	127	2	98	6.959	0.434
	Average	5.8	37.8	122	10	91	6.628	0.411
203.7	5	0.7	30.75	40	6	85	17.462	1.089
203.7	5	1.1	23.25	62	12	83	12.604	0.786
203.7	5	1.4	25.2	43	12	72	6.815	0.425
203.7	5	1.8	29.0	62	12	81	10.022	0.625
203.7	5	2.1	40.75	49	4	92	2.453	0.153
	Average	2.8	29.8	51	9	83	9.871	0.616
325.9	8	3.9	16.0	98	10	90	7.120	0.444
325.9	8	11.3	22.25	33	11	67	0.850	0.053
325.9	8	8.4	21.4	60	10	83	2.485	0.155
325.9	8	15.2	22.7	42	4	90	2.518	0.157
325.9	8	21.5	49.4	102	29	72	3.287	0.205
325.9	8	13.0	20.7	299	33	89	8.274	0.516
325.9	8	10.5	17.5	93	40	57	1.716	0.107
	Average	12.0	24.3	104	20	81	3.75	0.234
407.4	10	9.5	17.75	70	19	73	2.325	0.145
407.4	10	10.2	18.6	49	17	65	1.427	0.089
	Average	9.9	18.2	60	18	70	1.876	0.117

TABLE D-2. DUAL MEDIA FILTER DATA SUMMARY SUPPLEMENT
RAW BLAST FURNACE WASTEWATER FEED

Run no.	Flow gpm/ft ²	Turbidity (FTU)		pH		Temp (°C)	
		IN	OUT	IN	OUT	IN	OUT
1	5	40	3.5	6.6	6.6	23	23
2	5	50	5.0	7.8	7.8	23	23
3	10	44	20.0	7.8	7.8	22	22
4	10	45	15	7.7	7.7	22	22
5	8	28	15	7.3	7.3	18	18
6	5	35	5	7.4	7.4	21	21
7	5	-	-	-	-	-	-
8	5	33	20	7.5	7.5	19	19
9	8	37	7	7.5	7.5	22	22
10	8	-	-	7.4	7.4	21	21
11	8	-	-	7.4	7.4	23	23
12	8	-	-	7.3	7.3	24	24
13	8	-	-	-	-	-	-
14	8	23	17	7.0	7.0	-	-
15	8	33	9	6.9	6.9	-	-
16	3	40	1.1	6.9	6.7	-	-
17	3	40	6.0	7.3	7.3	-	-
18	3	40	5.1	7.3	7.3	-	-
19	5	45	4.5	7.0	6.8	-	-
AVERAGE		38	9.5	-	-	21.6	21.6

Note: $\text{gpm/ft}^2 \times 40.74 = \text{L/min/m}^2$

TABLE D-3. DUAL MEDIA FILTER - DATA SUMMARY
CHEMICALLY PRETREATED BLAST FURNACE WASTEWATER FEED

Filtration rate l/min/m ²	Run no.	Headloss meters	Run time hrs	Suspended solids, mg/l		Percent solids removal	Solids loading	
				Inf.	Eff.		kg SS/m ² /m H.L.	lb SS/ft ² /ft H.L.
122.2	25	2.1	30.25	25	15	40	1.04	0.065
203.7	6	9.5	7.66	23	6	74	0.16	0.010
203.7	7	4.9	23.58	28	15	46	0.76	0.047
203.7	8	12.0	22.75	27	3	89	0.56	0.035
203.7	9	7.7	25.4	20	7	65	0.44	0.028
203.7	10	5.6	18.5	28	*	-	-	-
203.7	11	9.1	30.0	22	13	41	0.43	0.027
203.7	12	9.1	30.0	21	9	57	0.51	0.032
203.7	13	7.7	25.4	25	8	68	0.47	0.029
203.7	14	7.0	23.1	26	10	62	0.75	0.047
203.7	15	4.2	13.9	48	23	52	1.64	0.102
203.7	16	3.1	30.0	16	3	81	0.42	0.026
203.7	17	10.6	34.7	26	8	69	0.50	0.031
203.7	18	11.3	37.0	21	18	14	0.06	0.004
203.7	19	10.6	34.7	25	2	92	0.61	0.038
203.7	20	14.1	46.2	33	4	88	0.56	0.035
203.7	21	7.0	23.1	20	6	70	0.56	0.035
203.7	22	5.6	18.5	23	5	78	0.90	0.056
203.7	23	23.2	23.2	31	18	42	-	-
203.7	24	4.2	13.9	20	14	30	0.42	0.026
	Average	8.4	27.8	27	10	66	0.55	0.036
325.9	1	9.5	31.2	42	37	12	0.35	0.022
325.9	2	9.3	32.3	32	*	-	-	-
305.6	3	8.1	26.6	21	10	52	0.67	0.042
305.6	4	5.6	18.5	26	25	4	0.03	0.002
305.6	5	15.5	50.8	24	12	50	0.34	0.021
	Average	9.6	31.8	28	21	30	0.35	0.022

* Sampling Difficulty

TABLE D-4. DUAL MEDIA FILTER DATA SUMMARY SUPPLEMENT
CHEMICALLY PRETREATED BLAST FURNACE BLOWDOWN

Run no.	Flow gpm/ft ²	Turbidity (FTU)		pH		Temp (°C)	
		IN	OUT	IN	OUT	IN	OUT
1	8	8.0	6.0	7.3	7.3	24	24
2	8	9.2	6.0	7.3	7.3	22	22
3	7.5	7.1	3.2	7.3	7.3	24	24
4	7.5	7.0	2.0	7.3	7.3	24	24
5	7.5	17	5.1	7.3	7.3	23	23
6	5	9	2.5	7.3	7.2	24	24
7	5	13	4.5	7.3	7.2		
8	5	17	2.0	7.3	7.2	24	24
9	5	9	3.0	7.2	7.1	24	25
10	5	8	2.4	7.2	7.0	25	25
11	5	15	3.0	7.2	7.2	21	21
12	5	15	3.2	7.2	7.1	23	23
13	5	18	4.0	7.3	7.2	22	23
14	5	18	4.5	7.2	7.2	21	22
15	5	26	5.0	7.3	7.2		
16	5	16	6.5	7.0	7.0	26	26
17	5	16	5.0	7.0	7.0	27	28
18	5	12	4.4	7.0	7.0	28	29
19	5	14	6.0	6.8	6.7	24	25
20	5	16	6.0	6.7	6.7	25	25
21	5	12	5.2	6.9	6.9		
22	5	12	3.0	6.9	6.9	26	26
23	5	14	7	6.9	6.9	22	22
24	5	4.5	1.2	6.8	6.8	22	22
25	3	6.5	1.2	6.9	6.9	23	23
AVERAGE		12.2	4.1	-	-	24.9	25.2

Note: $\text{gpm/ft}^2 \times 40.74 = \text{l/min/m}^2$

TABLE E-1. MAGNETIC FILTER-DATA SUMMARY
RAW BLAST FURNACE WASTEWATER FEED

Filtration rate $\frac{\text{L}}{\text{min}/\text{m}^2}$	Run no.	Headloss meters	Run time hrs	Suspended solids, mg/L		Percent solids removal	Solids loading	
				Inf.	Eff.		kg SS/m ² /m H.L.	lb SS/ft ² /ft H.L.
534	13	3	2.0	97	8	92	0.88	.055
1589	39	11	1.5	66	42	36	0.30	.019
1589	39	9	1.5	60	29	52	0.48	.030
1589	39	13	4.2	33	21	36	0.50	.031
1589	39	10	13	*	*	-	-	-
1589	39	15	4.9	128	46	64	1.31	.082
1589	39	16	2.2	128	32	75	1.30	.081
1589	39	9	0.75	160	22	86	1.07	.067
Average		12	2.3	96	32	58	0.83	.052
2648	65	20	6.5	54	31	43	1.17	.073
2648	65	10	1.5	49	34	31	0.36	.023
2648	65	16	2.2	49	27	45	0.50	.031
2648	65	*	*	-	-	-	-	-
2648	65	16	1.8	38	39	-	-	-
Average		15	3.4	51	31	40	0.57	.042

* Sampling Difficulties

TABLE E-2. MAGNETIC FILTER - DATA SUMMARY SUPPLEMENT
RAW BLAST FURNACE WASTEWATER FEED

Run no.	Flow [*] gpm/ft ²	Turbidity (FTU)		pH		Temp (°C)	
		IN	OUT	IN	OUT	IN	OUT
1	65	39	32	6.5	6.5	21	21
2	65	55	35	6.7	6.7	24	24
3	65	44	30	7.8	7.8	23	23
4	39	55	40	7.8	7.8	23	23
5	39	43	30	7.8	7.8	22	22
6	65	33	28	7.5	7.5	21	21
7	65	32	25	7.5	7.5	23	23
8	65	36	27	7.6	7.6	22	22
9	39	30	20	7.3	7.2	22	22
10	39	-	-	7.3	7.3	-	-
11	39	36	30	7.3	7.3	-	-
12	39	33	23	7.5	7.5	-	-
13	39	31	18	7.4	7.4	-	-
14	13	59	9	7.1	7.1	-	-
AVERAGE		40	27	-	-	22	22

* gpm/ft² x 40.74 = l/min/m²

TABLE E-3. MAGNETIC FILTER - DATA SUMMARY
CHEMICALLY PRETREATED BLAST FURNACE WASTEWATER FEED

Filtration rate l/min/m ²	gpm/ft ²	Run No.	Headloss		Run time hrs	Suspended solids, mg/L		Percent solids removal	Solids loading	
			meters	feet		Inf.	Eff.		kg SS/m ² /m H.L.	lb SS/ft ² /ft H.L.
794	19.5	5	4.2	13.8	5.5	25	21	16	0.257	.016
1589	39	1	2.8	9.2	3.0	30	10	67	2.037	.127
1589	39	2	2.1	6.9	1.5	33	12	64	1.427	.089
		Average	2.4	8.0	2.2	31	11	65	1.732	.108
2648	65	3	6.3	20.8	4.8	25	-	-	-	-
2648	65	4	6.3	20.8	3.4	25	24	-	-	-
		Average	6.3	20.8	4.1	25	24	-	-	-
(With Magnetite)										
1589	39	6	4.9	16.2	1.5	608	581	4	0.786	.049

TABLE E-4. MAGNETIC FILTER - DATA SUMMARY SUPPLEMENT
CHEMICALLY PRETREATED BLAST FURNACE BLOWDOWN

Run no.	Flow [*] gpm/ft ²	Turbidity (FTU)		pH		Temp (°C)	
		IN	OUT	IN	OUT	IN	OUT
1	39	17	8	7.2	7.2	22	23
2	39	17	11	6.9	6.9	-	-
3	65	6.0	4.9	6.9	6.9	23	23
4	65	6.3	5.2	6.9	6.9	23	23
5	19.5	5.3	3.3	6.8	6.8	22	23
AVERAGE		10.3	6.5	-	-	23	23
(WITH MAGNETITE)							
6	39	25	18	6.9	6.9	-	-

* gpm/ft² x 40.74 = l/min/m²

TABLE F-1. BLAST FURNACE REVERSE OSMOSIS FEEDWATER-PHASE 1

Parameter	Date		
	3/2	3/3	3/4
pH	7.64	7.60	7.70
SS, mg/l	57	10	25
VSS, mg/l	0	5	6
Phenol, mg/l	0.012	0.015	0.013
NH ₃ as N, mg/l	32.2	30.7	41.8
NO ₃ as N, mg/l	0.11	0.10	0.07
TKN as N, mg/l	34.2	33.7	43.2
TP as P, mg/l	0.06	0.03	<0.01
S, mg/l	<0.2	<0.2	<0.2
CN, mg/l	0.07	0.05	0.05
CN _A , mg/l	0.07		
F, mg/l	9.7	10.9	10.3
TOC, mg/l	6	2	2
Cond., μ mhos/cm	1500	2000	2075
Color, APHA units	0/5	0/5	0
TDS, mg/l	1119	1380	1396
Alk., mg/l as CaCO ₃	275	285	280
Cl, mg/l	380	400	385
SO ₄ , mg/l	155	177	220
Na, mg/l	16	18	17
Ca, mg/l	56	58	53
K, mg/l	22	25	25
Mg, mg/l	50	60	59
Mn, mg/l	3.3	3.4	2.6
Fe, mg/l	6.0	2.3	1.4

TABLE F-2. BLAST FURNACE REVERSE OSMOSIS FEEDWATER - PHASE 2

Parameter	Date												
	3/5	3/6	3/7	3/8	3/9	3/11	3/14	3/16	3/18	3/19	3/21	3/22	3/23
pH	7.68	7.85	7.40	7.60	7.60	7.90	6.50	7.10	6.70	6.60	6.70	6.70	6.50
SS, mg/l	12	6	15	3	7	19	13	8	6	12	9	9	8
VSS, mg/l	4	4	6	1						3	3	3	3
Phenol, mg/l	0.014	0.028	0.011	0.010	0.008	0.014	0.151			0.176	0.025	0.015	0.015
NH ₃ as N, mg/l	30.4	48.5	43.0	43.4	44.8	36.5	46.0	47.0	37.0	33.5	4.5	26.0	27.0
HCl as N, mg/l	0.95	0.10			0.07								
TKN as N, mg/l	35.0	50.2					43.4						
TP as P, mg/l	<0.01		0.01		<0.01					3.6	26.4	27.0	
S, mg/l	<0.2	<0.2					<0.2			<0.2	<0.2	<0.2	<0.2
CN, mg/l	0.04	0.02	0.02	<0.01	<0.01	<0.01	0.32	0.04		<0.01	<0.01	<0.01	<0.01
CN _A , mg/l	0.03	0.02	0.02	<0.01	<0.01	<0.01	0.27	0.04		<0.01	<0.01	<0.01	<0.01
F, mg/l	10.8	8.0		8.58	6.90	9.9	7.8			3.7	5.2	6.6	
TOC, mg/l	1	2	2	1	1	5	2		5	4	1	2	1
BOD ₅ , mg/l					<2								
Cond. umhos/cm	2000	2240	2000	2250	2350	1950		2250		1950	1970	2100	
Color APHA units	0/5	0/5	0/5	0			0						
TDS, mg/l	1331	1418	1376	1419	1648	1210	1353	1340	1379	1305	1320	1364	1405
Alk., mg/l	270		330		250								
Cl, mg/l	380	413	405		520		470		555		220	200	525
SO ₄ , mg/l	195		165		160		190						
Na, mg/l	16		16		16		27						
Ca, mg/l	51		120		120		133						
K, mg/l	23		26		24		24						
Mg, mg/l	52		60		5.2		63						
Mn, mg/l	3.1		3.2		4.7		5.2						
Fe, mg/l	1.2		0.5		0.4		0.6						

TABLE F-3. BLAST FURNACE REVERSE OSMOSIS OPERATING DATA - PHASE I

Run #	Date	Time	Run time, hrs.	Phase time, hrs.	Feed press., psi	Temp., °C	Press., Corr. ^a	Temp., Corr. ^b	Product flow gpm	Product flow gpd ^c
1	2/28	1645			215	24.7	1.86	1.01	4.50	6483
		1745	1.0	1.0	225	25.0	1.78	1.00	4.27	6144
		2230	5.75	5.75	235	25.0	1.70	1.00	4.00	5760
	3/1	1750	5.75	5.75	210	23.9	1.90	1.03	4.72	6803
		2145	9.83	9.83	215	24.2	1.86	1.02	4.39	6315
3/2		1145	15.08	15.08	225	26.7	1.78	0.95	4.06	5849
		1530	18.83	18.83	225	26.4	1.78	0.96	4.01	5774
		1850	22.16	22.16	235	25.8	1.70	0.98	3.99	5739
		2145	24.83	24.83	240	25.5	1.67	0.98	3.93	5666
3/3		1445	34.83	34.83	235	26.7	1.70	0.95	3.89	5600
		1745	37.83	37.83	240	26.7	1.67	0.95	3.73	5369
		2225	42.50	42.50	245	26.9	1.63	0.94	3.62	5216
3/4		1300	45.83	45.83	250	25.8	1.60	0.98	3.67	5283
		1445	47.58	47.58	255	25.8	1.57	0.98	3.60	5179
		1445	47.58	47.58	235	25.0	1.70	1.00	4.08	5882

a. Pressure correction = 400/P actual

b. Temperature correction = 1.03 (250 - T)

c. Flow at 250 °C, 400 psi

NOTE: psi x 703.1 = kg/m²
gpm x 3.785 = l/min
gpd x 3.785 = l/day

TABLE F-4. BLAST FURNACE REVERSE OSMOSIS OPERATING DATA - PHASE 2

Run #	Date	Time	Run time, hrs.	Phase time, hrs.	Feed press., psi	Temp., °C	Press., Corr. a	Temp., Corr. b	Product flow gpm	Product flow gpd
1	3/5	1330	2.17	2.17	235	26.7	1.70	0.95	3.89	5600
		1540			255	29.7	1.57	0.87	3.21	4617
2	3/5	1815	5.0	2.17	212	27.8	1.89	0.92	4.17	6007
		2315		7.17	215	31.1	1.86	0.83	3.73	5367
3	3/6	0800	13.75	15.92	245	31.1	1.63	0.83	3.00	4317
		1200		15.92	220	27.8	1.82	0.92	4.02	5788
4	3/7	1325	3.75	19.67	245	26.9	1.63	0.94	3.70	5327
		1710		23.42	255	30.0	1.57	0.86	3.25	4676
5	3/8	2200	8.58	28.25	258	30.0	1.55	0.86	3.08	4429
		0800		38.25	258	30.0	1.55	0.86	3.08	4429
6	3/9	1210	22.75	42.42	275	28.3	1.45	0.91	3.16	4555
		1630		46.75	265	30.0	1.51	0.86	3.06	4406
7	3/10	2100	31.58	51.25	275	31.1	1.45	0.83	2.97	4284
		0730		61.75	270	31.1	1.48	0.83	2.91	4185
8	3/10	0910	43.75	63.42	277	31.1	1.44	0.83	2.77	3992
		1200		66.25	280	30.6	1.43	0.85	2.91	4189
9	3/10	1640	51.25	70.92	280	30.0	1.43	0.86	2.96	4259
		2205		76.34	275	31.1	1.45	0.83	2.91	4196
10	3/10	0925	68.00	87.67	278	27.8	1.44	0.92	3.11	4485
		1350		92.09	285	28.9	1.40	0.89	3.00	4324
11	3/10	1940	92.09	92.09	255	23.3	1.57	1.05	3.95	5695

continued

TABLE F-4. (continued)

Run #	Date	Time	time, hrs.	Phase time, hrs.	Feed press., psi	Temp., °C	Press., Corr.a	Temp., Corr.b	Product flow gpm	Product flow gpd
5	3/11	1415	7.0	99.09	260	26.7	1.54	0.95	3.51	5061
		1700	9.75	101.84	274	27.2	1.46	0.94	3.28	4724
		2130	14.25	106.34	280	27.2	1.43	0.94	3.14	4527
6	3/12	0855		106.34	280	22.8	1.43	1.07	3.66	5272
		1025	1.50	107.84	273	27.8	1.46	0.92	3.31	4762
		1330	4.58	110.92	282	28.3	1.42	0.91	3.15	4535
		1420	5.41	111.75	272	27.8	1.47	0.92	3.45	4974
		1435	5.66	112.00	252	27.8	1.59	0.92	3.44	4948
		1625	7.49	113.83	260	28.9	1.54	0.89	3.22	4641
		1930	10.57	116.91	278	28.9	1.44	0.89	3.08	4433
		2255	13.99	120.33	280	28.9	1.43	0.89	3.06	4401
		0820	24.41	130.75	285	29.4	1.40	0.88	2.89	4165
		1150	27.91	134.25	290	30.0	1.38	0.86	2.85	4112
7	3/13	1400	30.08	136.42	288	31.1	1.39	0.83	2.78	4007
		1815		136.42	245	26.7	1.63	0.95	3.73	5371
		1825	0.17	136.59	245	27.8	1.63	0.92	3.46	4981
		1950	1.59	138.01	268	30.0	1.49	0.86	3.09	4449
		2300	4.76	141.18	273	30.0	1.46	0.86	2.97	4277
8	3/14	0830	14.26	150.68	300	27.8	1.33	0.92	2.82	4068
		1230		150.68	265	26.1	1.51	0.97	3.51	5048
		1450	2.33	153.01	275	27.2	1.45	0.94	3.13	4511
9	3/15	1815	5.75	156.43	292	27.2	1.37	0.94	3.01	4341
		1100		156.43	290	23.3	1.38	1.05	3.48	5008
		1200	1.00	157.43	292	26.7	1.37	0.95	3.26	4694
		1630	5.50	161.93	282	27.8	1.42	0.92	3.07	4422

continued

TABLE F-4. (continued)

Run #	Date	Time	Run time, hrs.	Phase time, hrs.	Feed press., psi	Temp., °C	Press., Corra	Temp., Corrb	Product flow gpm	Product flow gpd ^c
9	3/15	1930	8.50	164.93	290	27.8	1.38	0.92	3.05	4391
		2215	11.25	167.68	290	28.3	1.38	0.91	2.94	4230
	3/16	0915	22.25	178.68	325	27.8	1.23	0.92	2.78	4000
		1230	25.50	181.93	400	27.8	1.00	0.92	1.84	2653
10	3/17	1130		181.93	325	25.0	1.23	1.00	3.01	4342
		1400	2.00	183.93	270	25.0	1.48	1.00	3.55	5120
		2130	9.17	191.10	285	30.6	1.40	0.85	2.86	4116
	3/18	1300	24.67	206.60	280	32.8	1.43	0.79	2.72	3923
	3/19	1315	48.92	230.85	275	33.8	1.45	0.77	2.63	3785
		1935	55.25	237.18	270	34.4	1.48	0.76	2.74	3953
	3/20	0815	67.92	249.85	274	32.2	1.46	0.81	2.89	4160
		1825	78.09	260.02	288	30.0	1.39	0.86	2.93	4227
	3/21	0845	92.42	274.35	285	30.6	1.40	0.85	2.86	4116
		1745	101.42	283.35	290	30.6	1.38	0.85	2.93	4213
		2020	104.00	285.93	275	30.6	1.45	0.85	2.90	4177
	3/22	0830	116.17	298.10	277	31.7	1.44	0.82	2.85	4098
		1610	123.84	305.77	277	35.0	1.44	0.74	2.63	3791
	3/23	0030	132.17	314.10	285	33.9	1.40	0.77	2.48	3574
		0920	141.00	322.93	295	33.6	1.36	0.77	2.62	3769
		1530	147.17	329.10	285	33.3	1.40	0.78	2.63	3791
		2305	154.59	336.52	290	33.3	1.38	0.78	2.59	3726
	3/24	0830	164.01	345.94	300	30.0	1.33	0.86	2.59	3726
		1700	172.51	354.44	315	30.0	1.27	0.86	2.63	3786

continued

TABLE F-4. (continued)

Run #	Date	Time	Run time, hrs.	Phase time, hrs.	press., psi	Temp., °C	Press., Corr ^a	Temp., Corr ^b	Product flow gpm	Product flow gpd ^c
11	3/25	1200		354.44	305	26.7	1.31	0.95	3.00	4315
		1930	7.50	361.94	315	26.7	1.27	0.95	2.78	4003
	3/26	1825	7.50	361.94	280	26.7	1.43	0.95	3.26	4700
		2230	7.50	361.94	270	26.7	1.48	0.95	3.38	4874

a. Pressure correction = $400 / P_{\text{actual}}$ b. Temperature correction = $1.03 (250 - T)$

c. Flow at 250 °C, 400 psi

NOTE: $\text{psi} \times 703.1 = \text{kg/m}^2$ $\text{gpm} \times 3.785 = \text{L/min}$ $\text{gpd} \times 3.785 = \text{L/day}$

TABLE F-5. BLAST FURNACE REVERSE OSMOSIS PRODUCT-PHASE 1

Parameter	Date		
	3/2	3/3	3/4
pH	5.35	5.60	5.65
SS, mg/l	2	1	5
VSS, mg/l	1	1	2
Phenol, mg/l	0.071	0.122	0.151
NH ₃ as N, mg/l	3.3	9.0	3.4
NO ₃ as N, mg/l	<0.01	<0.01	<0.01
TKN as N, mg/l	5.5	11.2	4.5
TP as P, mg/l	0.01	<0.01	<0.01
S, mg/l	<0.2	<0.2	<0.2
CN, mg/l	<0.01	<0.01	<0.01
F, mg/l	0.28	0.30	0.30
TOC, mg/l	1	<1	<1
Cond., μ mhos/cm	96	136	162
Color, APHA units	0	0	0
TDS, mg/l	15	41	9
Alk., mg/l as CaCO ₃	10	20	20
Cl, mg/l	15	20	20
SO ₄ , mg/l	<2	<2	<2
Na, mg/l	2.2	1.2	1.6
Ca, mg/l	1.6	1.8	2.8
K, mg/l	1.8	2.0	2.1
Mg, mg/l	1.2	2.0	3.0
Mn, mg/l	0.10	0.06	0.08
Fe, mg/l	1.1	0.4	0.3

TABLE F-6. BLAST FURNACE REVERSE OSMOSIS PRODUCT - PHASE 2

Parameter	Date														
	3/6	3/7	3/8	3/9	3/11	3/12	3/14	3/15	3/17	3/18	3/19	3/21	3/22	3/23	3/25
pH	5.95	6.30	6.60	5.90	6.70	5.50	6.00	7.30	8.20			5.80	7.10	6.30	6.80
SS, mg/l	5		3				6		1						
VSS, mg/l	4						2								
Phenol, mg/l	0.128	0.143	0.030			0.072	0.023		0.10			0.016	0.007	0.018	<0.001
NH ₃ as N, mg/l	17.3	30.4	21.4	25.5	21.5	19.0	28.0	23.3	15.0	14.0	12.3	16.5	15.5	20.0	19.8
NO ₃ as N, mg/l	<0.01		<0.01				0.03		0.04						
TKN as N, mg/l	18.6						24.0								
TP as P, mg/l	<0.01		<0.01				<0.02		0.02			13.0	16.5	13.0	19.6
S, mg/l	<0.2	<0.2										<0.2	<0.2	<0.2	<0.2
CN, mg/l	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01			<0.01	<0.01	<0.01	<0.01
CN _A , mg/l	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01			<0.01	<0.01	<0.01	<0.01
F, mg/l	0.8		0.66			0.73	0.65		0.41			0.32	0.61	0.68	0.78
TOC, mg/l	1	<1	<1	<1	<1	1	4		<2	<2	<2	<1	<1	4	9
BOD ₅ , mg/l			<2												
Cond., umhos/cm	320	570	430		380		530		340			250	540	360	350
Color APHA units	0	0	0						0			0			
TDS, mg/l	30	157	134	110	180	85	71	112	172	148	149	195	229	106	141
Alk., mg/l as CaCO ₃	30		45				54		36						
Cl, mg/l	60	120	75				78		64	61		64	91	25	65
SO ₄ , mg/l	<2		1				<1		<2						
Na, mg/l			4.2				5.4								
Ca, mg/l			8.7				2.2								
K, mg/l			6.9				5.4								
Mg, mg/l			8.3				1.7								
Mn, mg/l			0.40				0.18								
Fe, mg/l			0.3				0.1								

TABLE F-7. BLAST FURNACE REVERSE OSMOSIS BRINE - PHASE 2

Parameter	Date															
	3/6	3/7	3/8	3/9	3/11	3/12	3/13	3/14	3/15	3/17	3/18	3/19	3/21	3/22	3/23	3/25
pH	7.05	7.20	7.20	7.30	7.30	7.10	7.20	7.50	7.50	6.70			6.50	6.70	6.50	7.40
SS, mg/l			26					21		29						
Phenol, mg/l	0.580	0.535	0.563			0.595	0.029		0.228	0.053		0.659	0.023	0.022	0.011	0.016
NH ₃ as N, mg/l	109.0	130.5	117.0	136.0	100.0	159.0	115.0	133.0	172.0	60.0	152.0	87.0	90.0	123.0	89	73
NO ₃ as N, mg/l	0.31		0.27	0.55		0.60		0.44		0.54						
TKN as N, mg/l	110.2							116.5					77.0	120	77.0	66.6
TP as P, mg/l	0.07		0.01					<0.02		0.06			<0.2	<0.2	<0.2	<0.2
S, mg/l	<0.2	<0.2	<0.01	<0.01	<0.01	<0.01	<0.2		0.38	<0.01			0.11	<0.01	0.11	<0.01
CN, mg/l	0.09	0.04	<0.01	<0.01	<0.01	<0.01	0.51	0.95	0.38	<0.01			0.08	<0.01	0.08	<0.01
CH _A , mg/l	0.04	0.01	<0.01	<0.01	<0.01	<0.01	0.51	0.95	0.38	<0.01			9.37	35.2	9.37	11.44
F, mg/l	13.4			13.0		6.6	21.7	25.1	21.3	22.6			16	10	16	10
TOC, mg/l	13	13	12	13	17	13	20	24	14	27	10	10	16	10	16	10
BOD ₅ , mg/l			7													
Cond., umhos/cm	6800	8500	6700		7000		7500	7800	6300	7500			8000	9800	8000	6000
Color APHA units	0/5	5	0/5					5	5	0						
TDS, mg/l	4951	6119	4794		4840	4540	5262	5695	4665	5644	5341	5616	5500	7655	5500	4310
Alk., mg/l as CaCO ₃			593					588		286			250	860	250	885
Cl, mg/l		2325						2130	1800	2285	2110		2325	3050	2325	1350
SO ₄ , mg/l	660		675							820						
Na, mg/l			50					100								
Ca, mg/l			560					94								
K, mg/l			75					1090								
Mg, mg/l			240					290								
Mn, mg/l			17					24								
Fe, mg/l			0.8					0.5								

TABLE G-1. OPERATIONAL DATA FOR THE FILTERED WASTEWATER OZONATION STUDY

Date	1/10/79	1/11/79	1/12/79	1/12/79	1/12/79	1/13/79	1/15/79	1/16/79
Run number	1	2	3	4	5	6	7	8
Influent flow, gpm	5.0	5.0	5.0	5.0	2.5	1.25	1.25	1.25
Influent temperature °C	22.2	23.0	23.9	25.0	25.0	27.0	19.5	21.3
Applied O ₃ , mg/l	269	323	500	1,500	500	1,054	460	695
Utilized O ₃ , mg/l	176	286	60	60	120	363	359	275
Contact time, minutes	60	60	60	60	120	240	240	240
Ozonator power setting, %	85	85	90	95	85	85	80	80
O ₃ gas temperature °C	17.7	22.0	41.1	41.1	29.4	22.2	14.7	15.0
O ₃ gas pressure, psi	15	15	15	15	15	15	15	15
Atmospheric pressure, mm Hg	757.3	758.1	751.5	750.3	747.5	740.6	751.5	750.7
O ₃ gas concentration, mg/l	20.32	20.32	35.82	69.49	20.32	20.32	13.23	18.42
Gas flow, scfm								
Contactor No. 1	4.0	3.8	4.0	7.0	4.3	4.3	2.9	4.43
Contactor No. 2	3.0	3.5	3.5	5.0	2.4	2.4	1.6	2.44
Contactor No. 3	1.8	3.3	2.5	3.5	2.0	2.0	1.3	1.99
Off-gas O ₃ concentration, mg/l								
Contactor No. 1	10.47	6.37		14.6		14.24	2.83	6.21
Contactor No. 2	5.35	0.00		16.1		13.13	2.11	7.84
Contactor No. 3	1.77	0.00		18.9		11.40	4.58	7.76
Effluent NH ₃ concentration, mg/l	34	46	40.5	40.5	33.0	3.1	29.0	13.7

(continued)

Note: gpm x 3.785 = l/min
 psi x 703.1 = kg/m²
 scfm x 28.32 = l/min

TABLE G-1. (continued)

Date	1/17/79	1/17/79	1/18/79	1/19/79	1/19/79	1/20/79	1/21/79	1/22/79
Run number	9	10	11	12	13	14	15	16
Influent flow, gpm	2.5	2.5	2.5	2.5	2.5	1.75	2.5	2.5
Influent temperature °C	25.0	23.0	25.6	26.7	23.6	25.6	25.6	32.8
Applied O ₃ , mg/l	1,000	1,205		735	781	1,074	0	739
Utilized O ₃ , mg/l		364		329	410	430	0	612
Contact time, minutes	120	120	120	120	120	171	120	120
Ozonator power setting, %	85	90	90	90	95	90	0	96
O ₃ gas temperature, °C	22.2	34.7	35.3	32.2	42.2	37.2		42.8
O ₃ gas pressure, psi	15	15	15	15	15	15	15	15
Atmospheric pressure, mm Hg	745.8	747.1	753.8	743.9	744.2	735.5		744.0
O ₃ gas concentration, mg/l	20.32	26.91		13.7	15.3	13.95	0	16.25
Gas flow, scfm								
Contactor No. 1	8.85	7.5	8.8	8.8	8.7	8.8	9.0	8.7
Contactor No. 2	4.88	4.1	4.9	4.9	4.8	4.9	5.0	4.8
Contactor No. 3	3.96	3.3	3.9	3.9	3.9	3.9	4.0	3.9
Off gas O ₃ concentration, mg/l								
Contactor No. 1		8.9		9.6	9.3	9.3	0	3.41
Contactor No. 2		8.8		5.7	6.1	6.1	0	1.90
Contactor No. 3		13.2		3.9	3.5	4.1	0	1.00
Effluent NH ₃ concentration, mg/l	21	9.4	11.0	13.0	11.6	3.4	32.0	10

Note: gpm x 3.785 = l/min
 psi x 703.1 = kg/m²
 scfm x 28.32 = l/min

TABLE G-2. OPERATIONAL DATA FOR THE RO BRINE OZONATION STUDY

Date	3/7	3/8	3/9	3/10	3/11	3/12	3/13	3/14	3/15
Run number	1	2	3	4	5	6	7	8	9
Influent flow, gpm	2.5	2.5	1.25	1.25	1.25	1.0	1.0	0.75	0.75
Influent temperature °C	25.8	27.4	29.9	25.6	21.8	26.8	27.6	25.6	24.4
Applied O ₃ , mg/l	991	738	1176	0	1319	632	1704	1877	3000
Utilized O ₃ , mg/l	760	470	537	0	525	169	665	747	1465
Contact time, minutes	120	120	240	240	240	300	300	400	400
Ozonator power setting %	85	85	85	0	85	80	85	85	85
O ₃ gas temperature °C	25	27.5	23.9	15.6	24.7	15.9	26.9	26.6	26.8
O ₃ gas pressure, psi	15.0	15.0	15.0	15.0	15.0	15.0	15.0	15.0	15.0
Atmospheric pressure, mm Hg	738.5	741.6	741.2	745.2	745.0	746.9	740.4	745.5	757.3
O ₃ gas concentration, mg/l	20.32	20.32	20.32	0	20.32	15.43	20.32	20.32	20.32
Gas flow, scfm									
Contactor No. 1	7.0	5.0	3.0	3.0	5.0	3.3	4.0	2.9	5.8
Contactor No. 2	5.0	4.0	2.7	2.7	4.0	2.8	3.2	2.4	4.8
Contactor No. 3	4.4	3.3	2.5	2.5	3.3	2.6	2.7	2.1	4.2
Off-gas O ₃ concentration, mg/l									
Contactor No. 1	3.52	7.08	11.12	0	12.01	6.54	11.64	13.97	7.28
Contactor No. 2	4.22	6.77	14.58	0	6.90	6.57	15.63	14.98	11.81
Contactor No. 3	7.30	8.53	12.23	0	14.02	8.45	16.10	17.94	13.10
Effluent NH ₃ concentration mg/l	39.0	91.5	75.0	95.0	57.0	69.0	42	42	30

Note:

gpm x 3.785 = l/min.

psi x 703.1 = kg/m²

scfm x 28.32 = l/min.

TABLE H-1. ACTIVATED CARBON DATA SUMMARY

Date	Cum. Flow gal.	Cum. Time hrs:min.	Cl ₂ conc., mg/l.		NH ₃ comp., mg/l.	CH ₄ comp., mg/l.		Phenol, comp., mg/l.	Sulfide, comp., mg/l.		Fluoride, comp., mg/l.		S.S., comp., mg/l.		pH - Grab		Temp., grab		TKN, comp., 10 ⁻³ mg/l.	
			Inf.	Eff.	Inf.	Inf.	Eff.	Inf.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.
2/2/79	126	0:40	28.4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/2/79	336	2:05	40.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/2/79	438	3:00	46.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/2/79	798	4:45	49.4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/5/79	1,278	6:15	21.77	1.77	2.21	2.21	0.48	1.0	<0.01	<0.01	0.014	0.216	<0.2	<0.2	12.5	12.5	16	6	7.15	6.85
2/6/79	6,307	26:45	21.27	4.43	1.32	1.32	0.67	0.65	<0.01	<0.01	0.011	0.083	<0.2	<0.2	10.9	12.2	16	9	7.3	6.8
2/7/79	9,116	37:30	9.0	-	-	-	-	-	<0.01	<0.01	0.004	0.360	<0.2	<0.2	11.0	10.0	16	8	7.3	6.8
2/8/79	11,416	47:15	130	-	-	-	-	-	<0.01	<0.01	0.004	0.360	<0.2	<0.2	11.0	10.0	16	8	7.3	6.8
2/9/79	11,416	47:15	106.3	12.4	1.32	1.32	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/9/79	12,469	51:45	106.3	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/9/79	13,756	57:15	106.3	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/10/79	15,429	66:15	106.3	-	-	-	10.3	13.8	<0.01	<0.01	0.008	0.345	<0.2	<0.2	12.4	12.6	-	44	6.3	6.6
2/10/79	15,789	67:45	61.1	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/10/79	17,376	74:35	59.3	-	-	-	5.2	6.7	<0.01	<0.01	0.009	0.045	<0.2	<0.2	13.4	13.4	25	23	7.0	6.6
2/11/79	21,443	94:35	57.6	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/12/79	24,038	113:05	97.4	30.9	4.5	3.4	7.4	10.8	0.11	0.08	0.012	0.03	<0.2	<0.2	14.2	13.4	11	7	7.7	6.7
2/12/79	24,083	113:35	97.4	-	-	1.77	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/13/79	26,299	130:20	25	21.9	3.0	1.9	-	-	<0.01	<0.01	0.004	0.116	<0.2	<0.2	19.2	18.7	33	9	7.3	6.8
2/13/79	26,894	136:00	39	8.8	2.7	1.3	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/14/79	28,729	145:50	40.5	33.2	24.4	1.8	6.95	8.15	<0.01	<0.01	0.013	0.123	<0.2	<0.2	13.6	14.6	35	14	7.1	6.6
2/14/79	29,107	147:35	70.4	48.7	8.4	3.5	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2/15/79	32,163	165:35	79.3	20.0	2.95	1.77	10.9	15.9	<0.01	<0.01	0.009	0.007	<0.2	<0.2	15.9	15.7	33	18	6.8	6.6
2/15/79	34,774	261:20	113	70.9	2.2	7.3	5.7	8.35	<0.01	<0.01	0.009	0.011	<0.2	<0.2	13.9	13.8	29	26	7.0	7.0
2/16/79	34,543	261:15	113	28	3.5	7.1	7.35	10.7	<0.01	0.06	<0.001	0.004	<0.2	<0.2	20.8	21.1	29	26	7.0	7.0
2/17/79	38,647	279:15	95.2	14	3.5	8.8	7.95	7.95	<0.01	<0.01	<0.005	<0.005	<0.2	<0.2	18.0	17.9	25	23	7.0	7.1
2/17/79	38,647	279:15	95.2	14	3.5	8.8	11.9	14.0	<0.01	<0.01	0.005	0.002	<0.2	<0.2	16.8	17.3	37	21	7.3	6.5
2/22/79	63,065	298:10	69	4.4	4.8	7.5	4.8	5.2	<0.01	<0.01	0.005	0.005	<0.2	<0.2	16.9	17.2	36	29	7.3	6.5

1. Composite sample over 24 hrs. or length of unit operation, whichever is smaller.

2. Last day of dual column operation, column #2 shutdown, column #1 continued.

3. All data from column #1. Column #2 data in separate table. All chlorine data as total available chlorine.

4. All effluent data (except Cl₂) from column #2 from 2/2-2/8. The remainder is from column #1.

GLOSSARY

Alkaline Chlorination Effluent - Effluent from alkaline chlorination of raw wastewater.

Alkaline Chlorinated RO Brine - Effluent from alkaline chlorination of reverse osmosis brine or concentrate.

Chemically Pretreated DMF Effluent - Effluent from dual media filtration of chemically treated, clarified raw wastewater.

Chemically Pretreated Magnetic Filter Effluent - Effluent from magnetic filtration of chemically treated, clarified raw wastewater.

Cyanide (CN_A) - Cyanide amenable to chlorination.

DMF Effluent - Effluent from dual media filtration of raw wastewater.

Magnetic Filter Effluent - Effluent from magnetic filtration of raw wastewater.

Ozonated Filter Effluent - Effluent from ozonation of dual media filtered raw wastewater.

Ozonated RO Brine - Effluent from ozonation of reverse osmosis brine or concentrate.

Raw Wastewater - Blast furnace gas scrubber system blowdown.

RO Brine - Concentrate produced by reverse osmosis treatment. Pretreatment for reverse osmosis consisted of chemical treatment with alum and polymer, clarification, and dual media filtration.

Total Cyanide (CN_T) - Complex cyanide and cyanide amenable to chlorination.